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# PREFEASIBILITY OF AMMONIUM PARATUNGSTATE PRODUCTION AT THE MOUNT PLEASANT TUNGSTEN MINE NEW BRUNSWICK

(Technical Assessment,
Flowsheet Preliminary Design
and Costing)

by

J. Dean Thibault, P.Eng.

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## DISCLAIMER

An *independent consultant* has performed a review of tungsten separation and extraction technology, based on published literature. The technology has been tested on *Mount Pleasant ores*, acknowledging the developers of the technology and/or previous studies done on the ore. The conclusion of this study has to rely on the accuracy of published literature and data derived from standard industrial testing procedures.

This study is to be used only for the purpose of assisting with the development of Mount Pleasant tungsten ores. Technological assessment and testing done under this study are not intended for commercial design or ownership of technology for commercial benefit.

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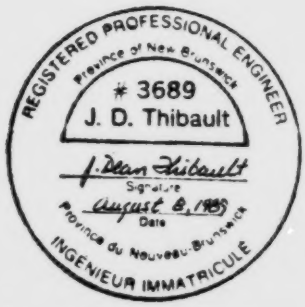

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## 1.0 EXECUTIVE SUMMARY

The closure of the Mount Pleasant Mine was the result of many factors, including the technical limitations of the process technology and a drop in the market value of tungsten concentrate. Final attempts to resolve product quality and production problems, using a grant from the Department of Regional Industrial Expansion (DRIE), were unsuccessful due to excess production costs. Since the mine closure, the feasibility of producing tin concentrates from an adjacent ore zone has been reviewed by Lac Minerals Ltd. while the Department of Natural Resources and Energy has been assessing development options for the tungsten ore body.

Under current market values, tungsten is still the main economic contributor to the value of the ore from the North Fire Tower Zone. In addition to tungsten, various development concepts for the production of by-products (such as calcium fluoride, molybdenum, bulk sulfides of copper, lead, and zinc, bismuth, and tin) have been identified. This study has focused on the development options applicable to tungsten products and has placed emphasis on the production of ammonium paratungstate (APT) from tungsten ores.

This report presents a pre-feasibility assessment of producing APT at Mount Pleasant based on a bench scale test program administered by the Department of Natural Resources and Energy. The major findings of this investigation can be summarized as follows:

- The proposed method of primary upgrading of the ores, by integrating grinding and gravity separation, is unique in the tungsten industry and should be more efficient than flotation technology. Process technology for the production of APT from gravity concentrates has been developed by various manufacturers but the development of process technology for Mount Pleasant ores should be ore specific because of the diverse concentration of impurities.

- The capital cost to construct a 2,235 tonne per year APT production plant within the existing facility at Mount Pleasant, is estimated to be approximately $28 million (January, 1989, Canadian funds). This estimate includes the purchase and installation costs of processing equipment and the indirect costs of grinding, gravity separation and the APT chemical plant. Operating cost estimates for processing of run-of-mine ore is equivalent to $13.50 per tonne of ore or $4.50 per kilogram of APT.

- The market for APT is quite competitive and the under-pricing of APT by the Chinese has been a major deterrent to the development of the North American tungsten industry. For a 10% return on investment over ten years (before taxes), the value of APT would have to be in excess of $150 per metric tonne unit (MTU) of tungsten trioxide contained in the concentrate, in order for the production of APT from Mt. Pleasant ores to be feasible. This is twice the published market value.

- The current price of tungsten powder and tungsten carbide has not been affected by the Chinese marketing practices. The production of tungsten carbide and hard metals from ore has projected revenues of approximately five times that of APT, with twice the capital investment required. The potential return on investment for a fully integrated production facility at Mount Pleasant is economically more attractive than the sole production of concentrate and/or APT.

- Since the production of APT is an intermediate processing step, the development of a fully integrated processing facility would require the development of APT processing technology. Additional processing technology for metal powder and carbide production would be considered standard industrial practice. Therefore, the technological development costs at Mount Pleasant would be equivalent to the development of APT production technology.

## 2.0 STUDY OBJECTIVES AND APPROACH

The primary objective of the present study is to assess the feasibility of APT production from ores at the Mount Pleasant Mine. In so doing, the assessment seeks to establish a development option for the commercialization of the Mount Pleasant tungsten ore zone. To assist in a better understanding of the process development requirements of Mount Pleasant ore, the study also includes an overview of test programs (previous to this study) administered by the New Brunswick Department of Natural Resources and Energy (DNRE) under the Federal-Provincial Mineral Development Agreement.

This study is comprised of preliminary engineering and production economics for a processing plant but does not include mining or ore handling considerations. Flowsheet design, equipment sizing and mass balance is derived from bench scale testing for the treatment of run-of-mine ore by grinding, gravity separation and chemical processing of APT. Preliminary order of magnitude costs are based on equipment supplier quotes for major equipment and factored equipment costs for a specific duty.

In order to address the above objectives, several sources of information are used to obtain technical and economic insights into the production of tungsten concentrates and APT. Sources of information include the following:

- o Technical literature search on processing of tungsten ores and concentrates.
- o Process Study Reports written by DNRE staff, documenting the results of the bench scale test program (Mineral Processing Investigations Program).
- o Mineralogical study of ores and concentrates by Witteck Development Ltd.
- o Review of tungsten industry economics, markets and developments as published by International Tungsten Symposiums, Roskill, Three River Associates, United States Bureau of Mines, and Energy Mines and Resources Canada policy sectors.
- o Discussions with industrial specialists and research groups including

CANMET, Hard Metal Research Inc., and Energy Mines and Resources Canada policy sectors.

- Correspondence with various equipment manufacturers pertaining to rod and ball mills, gravity separators, digesters, solvent extraction cells and crystallizers.

## 3.0 INTRODUCTION

Canada has the largest resource base of tungsten in North America, and is second in the world only to China. Canada is capable of producing tungsten concentrates from the wolframite ($FeWO_4$) ores of Mount Pleasant in New Brunswick and from the scheelite ($CaWO_4$) ores of Cantung in the Northwest Territories. Concentrates from each mine are exported to the United States and Europe for the manufacture of ammonium paratungstate and subsequent production of tungsten powder and/or tungsten carbide. At the present time both mines are shutdown due to low market prices.

Presently, the tungsten industry in central and western Canada is active in secondary manufacturing tungsten products by importing tungsten powder and tungsten carbide. It is interesting to note that in Canada, intermediate processing facilities for the conversion of tungsten concentrates to tungsten powder and tungsten carbide are nonexistent. To establish the reasons for this would require an additional assessment of the relationship between production, consumption, trade, and the price structure of the various tungsten products.

Tungsten powder and tungsten carbide are produced by the direct reduction of tungsten trioxide, which is produced by the calcination of APT. APT can be produced from scheelite and wolframite concentrates, however, because of the difference in chemical composition of the minerals, different processing technologies apply. For scheelite concentrates, hydrochloric leaching methods are commonly used to dissolve the tungsten and form a tungstic acid intermediate. APT is produced from the tungstic acid by the addition of an ammonia solution after the removal of impurities. Caustic leaching or the more selective soda pressure leach is used to produce APT from wolframite concentrates by the addition of an ammonia solution to a purified sodium tungstate.

In summary, the process technology for the production of tungsten powder and tungsten carbide concentrates is considered "adoptive technology", not requiring research and piloting to commercialize the technology. The competitiveness and

acceptance of process technology in the industry is related to the purity of APT. Marketing constraints are generally due to the purity of APT feedstock used to produce tungsten powder. Impurities such as silica, phosphate, arsenic, and molybdenum are detrimental to crystal formation of tungsten metal and hard metal alloys. Therefore, it is essential that impurities are removed in intermediate processing stages.

## 4.0 LIMITATIONS TO THE UPGRADING OF MOUNT PLEASANT ORES

During the operation of the Mount Pleasant Tungsten Mine (MPTM) by Billiton Canada Ltd., it was documented that the marketability of the concentrate and the efficiency of the concentrator were limited because of specific impurities in the ore. The nature of the ore was such that there was a complex variation in the mineralogy of the sections in the North Fire Tower Zone. Two major sections of the tungsten ore body have a significant variation in deleterious elements such as phosphorus and arsenic. Efforts to remove impurities at various stages in the processing of the ore had an adverse effect on tungsten recoveries.

### 4.1 RECOVERY OF TUNGSTEN

The concentrating circuit was unique to MPTM and was designed to recover 70 to 75% of the tungsten from run-of-mine ore. Technical audits on the commercial operation documented losses of tungsten resulting from desliming, flotation and magnetic separation unit operations. The actual overall operating recovery of tungsten was 50 to 55%, with the loss of metal due predominantly due to unrecoverable tungsten fines of less than 10 microns in size.

Over-grinding of the ore, which was a consequence of the auto-peb grind circuit and classification circuit design, generated a significant amount of slime or fine tungsten. In order to achieve a good liberation of tungsten grains, the net power requirements for a single stage grind has been shown to cause over-grinding. For a single stage grind of 80% less than 75 microns, the amount of tungsten less than 25 micron in the ground ore would vary from 45 to 60%. In order to achieve a desirable tungsten particle size distribution (between 15 and 75 micron)s for an optimum recovery, it was concluded that major design modifications would be required in the grinding and particle size classification circuits.

## 4.2 IMPURITIES IN THE CONCENTRATE

The concentrate produced by the MPTM flowsheet, contained significant levels of arsenic, phosphate, copper, lead and bismuth. These impurities limited the sale of concentrate for some specific end uses. The flowsheet was designed to upgrade the ore via flotation - magnetic separation. A ferric sulfate leach was used to remove impurities in the concentrate. If the impurity concentration in the ore increased or if the impurities occurred as major occlusions in the tungsten grains, the flowsheet was not capable of reducing impurity levels in the concentrate.

Arsenic was found to be the major impurity limiting the marketability of concentrate. The maximum limit for arsenic concentration in the concentrate was between 0.05 to 0.15% As, depending on the end use. Efforts to control ore grade and to maintain arsenic below 0.15% were unsuccessful. Lowering the arsenic content in flotation and magnetic separation circuits resulted in the excess loss of tungsten.

## 4.3 RECOVERY OF BY-PRODUCTS

The Mount Pleasant deposit has a diversified mineralogical composition and the interest in recovering elements such as molybdenum, tin, bismuth, indium, fluoride and base metals is based on the ore concentrations and the value of concentrates. The economic feasibility of a multi-mineral concentrator has not been studied in detail but the technical feasibility was studied by CANMET in 1970. It was recognized that the co-production of multi-minerals might offer competitive production costs since the ore has to be ground for tungsten recovery. However, the purity of each concentrate may be a major limitation to its economic feasibility. Subsequent testing by Billiton Canada Ltd., prior to commercial production and during the operation of the concentrator, resulted in the identification of the following concepts for the production of by-products:

*Molybdenum*: The recovery of molybdenum sulfide was implemented as a part of the MPTM flowsheet. The decision to recover molybdenum resulted in a significant production cost, as a result of chemical upgrading molybdenum concentrate (chemical upgrading of molybdenum concentrates was employed in the MPTM flowsheet to remove impurities such as arsenic, lead, etc.). In addition, a major limitation to the upgrading of molybdenum by flotation was the need for an ultra-fine grind, which made concentrate dewatering very difficult. As a result, the commercial production of molybdenum concentrate was not technically feasible or economically viable.

*Tin*: The recovery of tin was considered as a part of the MPTM flowsheet but the quantity of tin in the tungsten deposit was not sufficient to justify the capital investment. Tailings from the high gradient magnetic separation circuit varied in tin concentrations from 1 to 20% Sn and contained 50 to 55% of the tin in the run-of-mine ore. The tailings also contained high concentrates of arsenic (varying from 20 to 30% As). Because of high arsenic a detailed technical feasibility of tin recovery was not given much consideration.

*Base Metals*: The upgrading of base metal sulfides is technically quite attractive because of the separation from silica-gangue is considered relatively straight forward. Limitations to the technical viability of producing a bulk sulfide concentrate includes the low heat content of the concentrate ( high heat content is a prerequisite of base metal extraction by smelting)) and the high concentration of arsenic are both limitations to the technical feasibility of producing a bulk sulfide concentrate. Based on the operating data of the MPTM flowsheet, the bulk concentration contained 1-5% Cu, 1-5% Pb, 3-5% Bi, 10-25% As, 8-15% Zn, and trace indium. Bismuth was noted as a major contributor to the value of the bulk concentrate but the overall feasibility of producing marketable product(s) was not studied in detail.

## 5.0 DEVELOPMENT OPTIONS FOR THE MOUNT PLEASANT TUNGSTEN ORES

The decline in tungsten concentrate prices was a major deterrent to the continued economic success of the Mount Pleasant operation. In addition, various technical difficulties were identified that might limit the extraction efficiency of any future operation. Based on the knowledge gained with in-plant testing of flotation, gravity separation, magnetic separation and leaching unit operations, the following options have been identified by DNRE as possible development alternatives to improve the feasibility of Mount Pleasant.

### 5.1 PRODUCTION OF TUNGSTEN CONCENTRATE

It is evident from the testing done by MPTM and the demand for high purity tungsten products, that the production of tungsten concentrate at Mount Pleasant may not be technically feasible or economically viable. In addition to the flotation-magnetic separation flowsheet developed by Billiton for the production of tungsten concentrate, the use of a gravity-magnetic separation technology may also be possible. The complex and sometimes high levels of impurities in the ore would likely result in the production of some concentrates containing high levels of impurities. The efficiency of removing impurities by leaching with strong oxidants, such as acidic-ferric solutions and/or nitric acid, is questionable.

Based on commercial leaching of tungsten concentrates, it is possible that the impurities in the concentrate are "refractory", i.e. the grains of tungsten may enclose the impurity grains so that chemical removal is technically impossible. A detailed mineralogical study on the association of impurities in the concentrate would be considered academic. If the impurities could be leached from the concentrates, the revenue would not support the production costs of a chemical treatment plant.

## 5.2 BY-PRODUCT PRODUCTION

As indicated previously, the by-product production of other minerals may improve the overall economics of a processing facility. The value of various concentrates based on the composition of the ore was studied by DNRE. It was concluded that even at an all-time low concentrate value of $52/MTU $WO_3$, tungsten is the major economic contributor to the value of the ore body. This suggests that the North Fire Tower Zone of the Mount Pleasant deposit should produce tungsten as a primary product. Secondary products, which could contribute to revenue include:

- o Calcium fluoride concentrate, acid grade (97% $CaF_2$).
- o Bismuth concentrate.
- o Bulk sulfide concentrate (copper, lead, zinc).
- o Molybdenite concentrate (65% $MoS_2$).

## 5.3 ADD-ON CONCEPTS

In addition to the production of tungsten concentrates, tungsten powder production has been recognized as a potential business opportunity, increasing the revenue potential of the ore by about five times. Additional development concepts that have an individual add-on value in the range of 2 to 6 dollars per tonne of ore include:

- o Arsenic trioxide produced from arsenic concentrates.
- o The production of ammonium paratungstate as an intermediate in metal powder production.
- o Ferro-tungsten.
- o Molybdenum oxide produced from molybdenum sulfide concentrates.

## 5.4 SUMMARY OF DEVELOPMENT OPTIONS

Various development options have been identified in previous sections based on the value of the products that could be recovered from the Mount Pleasant ores. It is evident that a development option for tungsten will be the driving force for the commercialization of the North Fire Tower Zone. Since the production of high purity wolframite concentrate is not a viable alternative for Mount Pleasant ores, a development program encompassing a fully integrated processing facility for APT and subsequent production of tungsten powder and tungsten carbide, is considered technically more viable for the following reasons:

- o Chemical treatment of concentrate by dissolution of tungsten and purification of the solutions is a more efficient method of lowering impurity levels.
- o Upgrading the ores to lower grade concentrates will improve on recovery of tungsten.
- o Production of APT and added-value products will also increase the revenue potential of low-grade ores.

Several companies have established fully-integrated processing operations at tungsten mines. Operating tonnages are not greater than 2,000 mt/yr W with ore grades in the range of 0.2 to 2% $WO_3$. In most cases, a low-grade concentrate containing 20% $WO_3$ is used as feedstock to produce APT. Companies in the United States using this development approach to commercialize low-grade deposits include:

*Union Carbide*: Produces a gravity concentrate at the Pine Creek deposit in the Sierra Nevada in California and manufactures APT from concentrates at Bishop California.

*Amax*: Treats low-grade concentrate from the Northwest Territories at the Fort Madison, Iowa plant. In Iowa, APT and blue tungsten oxides are produced at a capacity of 1,000 tpa W. Cantung has just recently acquired the Fort Madison Plant.

*General Electric*: Operates the Springer mine in Nevada, which includes a mill and an APT plant. Tungsten metal and tungsten carbide are also produced by GE at a separate facility.

## 6.0 AMMONIUM PARATUNGSTATE AND THE TUNGSTEN INDUSTRY

Tungsten concentrate is sold on the world market at a minimum grade of 65% $WO_3$, which may be converted to APT or ferro-tungsten. In excess of 75% of the world demand for tungsten concentrates is used to produce tungsten powder by the calcination and hydrogen reduction of APT. Some of the tungsten powder is utilized in steel production, wires, lamp filaments, superalloys, and chemicals, while the bulk of it is carburized for the production of tungsten carbide.

### 6.1 APT TECHNOLOGY

The production of APT is still considered the centre of the tungsten industry because tungsten products are predominantly manufactured from APT. An alternative to APT production is the technology being developed to chlorinate low-grade concentrates, followed by the reaction of the chlorinated product with methane to produce tungsten carbide. This method of production would eliminate the APT stage. A commercial limitation to this technology is that the grain size of the tungsten carbide product cannot be controlled. Further research is required to demonstrate that this technology is competitive with APT technology.

The production of APT from wolframite ore of Mount Pleasant would included the chemical dissolution of concentrates using a sodium hydroxide leachant to produce a solution of sodium tungstate (commonly dissolved by digestion or pressure leaching), purification of the solution to remove arsenic, molybdenum, silica, etc. and reacting the sodium tungstate with ammonia to produce APT. Major by-products of the process include ferrous hydroxides and silicate precipitates. A proposed method of producing APT at the Mount Pleasant Mine (by combining gravity separation and chemical processing of APT) would eliminate the need to produce high-grade concentrates. Although this method of production is considered commercial, it will require further development to optimize and tailor the technology to Mount Pleasant ores.

## 6.2 PRODUCERS OF APT

The major producers of APT in North America are predominantly located in the United States and include: GTE, Union Carbide, General Electric, Amax, Li Tungsten, and Teledyne Wah Chang. The combined capacity is estimated to be 13,370 tonnes W/year. Each plant has a throughput ranging from 730 to 4,320 tonnes W/year with GTE being the largest producer. The average production rate for an APT producer is calculated to be 1,900 tonnes W/year, and most of the production is subsequently used to manufacture tungsten powder.

China is expanding from sole production of concentrates into downstream production of APT, which has resulted in a more competitive market for tungsten. At least twelve APT producers are known to be presently in operation in China. The price spread between concentrate and APT has varied over the last ten years and the price of APT from China has sometimes been lower than the concentrate market price. The smaller mines in North America producing concentrate have an operating cost of $120 to $160/MTU of $WO_3$ (Canadian funds). These operating costs are well in excess of the potential revenue that may be obtained from the sale of concentrates. APT production costs have not been reported in the literature, however, the ability to compete with the Chinese selling APT at prices lower than concentrate is a major deterrent to APT production.

## 6.3 PRODUCT SPECIFICATIONS

Consistency in product quality has been the key to establishing long term markets in North America. The chemical processors of APT for metal powder production have tight specifications on the concentrations of impurities, which is related to the hardness of a tungsten product. A recent change in specifications is due to improvements in fabrication and the quality of metal product. The limits placed on impurity concentrations in APT, used for the production of cemented carbide products, is in the range of 10 to 50 ppm.

## 7.0 BENCH SCALE TEST PROGRAM

A test program was designed and implemented by DNRE to determine the prefeasibility of gravity separation and the chemical treatment of gravity concentrate to produce APT. Gravity separation and grinding tests were done on ore collected as a bulk sample of plant feed during the operation of MPTM concentrator. As well, hydrometallurgical testing of the APT flowsheet was done on flotation concentrate obtained form the former MPTM operation. From the bench scale test data a preliminary flowsheet was proposed as a basis for process economics. Operating and capital costs were then projected in order to assess process economics.

### 7.1 SUMMARY OF TECHNICAL LITERATURE

A computerized literature search of the Chemical Abstract databases 1967-1986 and Minproc (Canmet's database for mineral processing technology) was based on processing technology for the treatment of wolframite and/or scheelite ores and concentrates. Secondary manufacturing of tungsten powder and tungsten carbide was not included in the search.

#### 7.1.1 FLOTATION

A number of papers were reviewed on the flotation of wolframite and the effective use of collectors. Collector chemistry and surface electric properties of the minerals have been studied in attempts to improve on wolframite selectivity. As a result of various laboratory studies, the major limitation to wolframite selectivity and recovery has been attributed to particle size and the variation in the manganese content of wolframite mineralization ($Fe,MnWO_4$).

Commercial installations of wolframite flotation circuits have not been common but a significant amount of testing has been done on the pilot scale. Bulk sulfides and slimes are removed prior to tungsten flotation to avoid over-consumption of flotation reagents. Overall flotation recoveries have been reported in the range of 71 to 80%, not including the desliming losses at less than 8 microns. Desliming may decrease the recovery of tungsten by 10% when the ore is ground to 60 to 80% -75 microns. The recovery of slime tungsten by column flotation has been tested on a pilot scale and is still awaiting commercialization.

Technology has been developed for the separation of bismuth, tin and/or molybdenum from wolframite ores by flotation, although the concentration of impurities in concentrates (which may limit market value) have not been reported. The flotation of fluorite, molybdenum and bismuth are common to pilot plant studies, however, commercial operations are not reported.

### 7.1.2 GRAVITY SEPARATION

Gravity separation techniques are commercially used for primary upgrading of tin and tungsten ores. The separation of tin from tungsten - gravity - concentrates is accomplished using high gradient magnetic separation. Jigs are commonly used on tungsten ores for the first phase of separation. Usually three stages of jigs are designed to recover about 58% of the tungsten after two stages of grinding. In the second phase of separation, gravity separation tables are commonly used in cleaning circuits for the production of wolframite grades up to 25% $WO_3$.

A drawback to gravity separation, as was the case for flotation, is the loss of fines. Gravity concentrator recoveries vary from 60 to 85%. Various methods or devices for the recovery of tungsten from fines and/or slimes have recently been developed for gravity separation but commercial efficiencies have not been published.

### 7.1.3 HYDROMETALLURGY

Most hydrometallurgical operations for treating tungsten concentrates are used to remove impurities or to produce APT. Some wolframite concentrates are commercially leached with nitric acid to dissolve phosphate, arsenic, and rare earth minerals. For scheelite concentrates, the dissolution of tungsten for APT manufacture is commonly done with hydrochloric acid followed by a water leach. Wolframite concentrate is dissolved using a caustic bake or digestion followed by a water leach for the commercial production of APT. Recent developments in pressure leaching of wolframite concentrates have improved the selectivity of the leach and have reduced sodium hydroxide consumption.

For the production of APT, technology used to purify sodium tungstate solutions has been centered around the selection of ideal cementation reactants for the precipitation of impurities in solution and operating conditions for effective cementation to limit the loss of tungsten with impurities. In addition to purification, the advantages of solvent extraction and ion exchange technology for the removal of sodium from solution have been compared. The use of solvent extraction and hydrolysis of the strip solutions still remain a common industrial practice.

In summary, the operating data from various concentrators and chemical processing plants have not been published. The limitations to concentrate upgrading and the production of APT appear to be associated with the removal of impurities. Recent developments in high gradient magnetic separation have improved the recovery of tungsten in primary upgrading, however, the need for chemical operations to remove impurities in the concentrate has been identified. The technological success of producing tungsten powder, carbide, and tungsten chemicals is related to the effective removal of impurities. The technology used to purify sodium tungstate solutions is considered to be the most effective method of dealing with ore impurities ( as apposed to leaching impurities from concentrates) and in most cases solution purification is proprietary.

## 7.2 GRAVITY SEPARATION TESTING

Bench scale and pilot gravity separation tests were performed by the Department of Natural Resources and Energy on ore collected from the Mount Pleasant Concentrator in 1986. Physical separation tests were done on a Mozley Mineral Separator to study grade-recovery relationships and the effects of grind on grade and recovery. Additional testing at the Nova Scotia Technical University assessed the grade-recovery relationships using pilot scale tables and spirals.

The bench scale lab program identified that the grinding time and the number of grinding stages was dependent on the recovery of tungsten. Kinetic data from a single-stage batch grinding test illustrates that a large amount of liberated tungsten (based on particle sizes less than 75 micron) is produced at relatively expedient rate. As the amount of liberated tungsten increases with time in the mill charge, the amount of fine tungsten (less than 25 microns) increases at a comparable rate. With a 20 minute grind about 40 to 50% of the tungsten is less than 25 microns. As a result, the data suggests that tungsten (in the liberated state) can be recovered from the charge at about 0.25 to 0.50 times the time required for complete liberation. Therefore, a multiple stage grinding circuit with an intermediate separation of tungsten would improve the tungsten recovery by reducing the concentration of fine tungsten.

The particle size analysis of the crushed ore has shown that 25% of the tungsten is less than 75 microns, suggesting that good liberation of the tungsten can be achieved by crushing alone. Piloting testing on the crushed ore using a jig has shown that a portion of the tungsten is recoverable at a grade of 3.88% and 9.28% $WO_3$ with a respective recovery of 29.9% and 16.5%. The grinding kinetics and the analysis of the crushed product have indicated that the ore has wide variations in tungsten grain size. If possible, the tungsten should be recovered at various stages in the comminution circuit in order to prevent overgrinding.

The optimum grade and recovery of tungsten from single stage grinding and gravity separation indicates that the liberation size is in the range of 80-85% -75

microns. A major limitation to a single stage grind is the loss of tungsten. Particle size analysis data confirms that the major loss of tungsten is due to the loss in the -25 microns fraction. In the tails, approximately 63% of the tungsten is less than 25 microns. Separation tests done at various stages of grinding (using a regrind on tails) show a relative improvement in the recovery of tungsten. Based on the grade-recovery data for a single stage grind a grade of 8% $WO_3$ at 60% recovery can be produced from a coarse grind ( 80% -30 mesh). Thus a regrind of both tailings and concentrates would be required to improve on grade and recovery.

Spiral tests also done on ground ore (80%, -75 microns) have shown that there is a fair amount of tungsten hang-up on the spiral, limiting the spiral to coarse feed. On material ground to 100% -150 microns, tests were conducted on a Wilfley separation table. An optimum separation can be achieved by splitting the feed into two size fractions, a plus and minus 75 microns. For the +75 microns material, a grade of 16% $WO_3$ with 75% to 80% recovery was achieved. For the -75 microns material, a grade of 15.6% was achieved at 58% recovery.

The test results from the gravity pilot testing have provided a basis for a preliminary design of a gravity circuit and are complementary to the gravity testing done by MPTM. Gravity separation can result in an overall improvement in recovery relative to flotation only if the grinding, classification and separation are integrated in a closed circuit configuration.

The comparison of the separation done on the Mozley mineral separator and various types of gravity separation pilot equipment suggests that direct scale-up from the Mozley separator is not appropriate. The use of the Mozley separator for gravity circuit design may require further assessment of scale-up factors. It is believed that the mechanical efficiency of the Mozley mineral separator is higher than the operation of a pilot table. Ideal operating conditions of the table based on water flow, stroke, etc., would have to be studied in order to quantify scale-up factors.

## 7.3 MINERALOGY OF MOUNT PLEASANT ORES AND CONCENTRATES

Detailed mineralogy was done by Wittech Development Ltd. in 1987 to identify ore liberation characteristics of various minerals, mineralogical associations, and to provide information for the development of a conceptual flowsheet by the analysis of gravity concentrate from the test program and flotation concentrate from the MPTM operation. The wide variation in mineral grain sizes and the diversified mineralogy of the ore suggested that the optimum separation circuit for tungsten would be complex. The mineralogical data has provided a basis for assessing the grinding and separation characteristics of the ore.

Wolframite liberation data suggests that the grain size varies from 45 to 900 microns. About 50% of the tungsten can be liberated at 150 microns. As indicated by the grinding and gravity separation testing, the wide range in grain size supports the concept of staged grinding and separation. In addition, the high liberation of tungsten in the gravity concentrate indicates that the ore is overground. It is anticipated that the recovery of tungsten can be improved if low-grade concentrates are produced since total liberation of tungsten is not required.

Wolframite occurs as an iron-rich tungsten oxide known as "ferberite" and as a manganese rich "huebnerite". The presence of huebnerite may account for the periodic high loss of tungsten from magnetic separation during the MPTM operation. Intergrowths of arsenopyrite, loellingite, molybdenite, chalcopyrite and sphalerite with wolframite grains may result in a concentrate having high impurity concentrations. These observations may be detrimental to the efficiency of concentrate production, and it is unknown how these mineralogical characteristics vary with the location in the orebody.

In addition to the recovery of wolframite, the limitations of by-product recovery may be determined partially by the mineralogical observations. The fine grains of bismuth and chalcopyrite suggest that physical upgrading of these minerals can result in limited recovery and/or a relatively impure concentrate. Similarly, the

rare earth minerals have been identified as phosphates that occur as minor occlusions in fluorite minerals. The recovery of rare earths may also be limited to chemical upgrading and the occlusions of phosphorus in the fluorite may be detrimental to the sale of high purity calcium fluoride. The recovery of fluorite may be considered relatively straightforward since the liberation is approximately 80% at 100 microns.

The difference in the mineralogical composition of flotation and gravity concentrates is directly related to the method of ore separation. A flotation concentrate has sulfides removed previous to wolframite flotation and the gravity concentrate will contain the minerals of high specific gravity. The gravity concentrate is composed mainly of arsenopyrite/loellingite, wolframite, galena and silicates; the flotation concentrate differs in the lead and arsenic compositions. Coarse grains of galena and arsenopyrite suggest that coarse gravity separation can have high recoveries of lead and arsenic.

## 7.4 APT HYDROMETALLURGICAL TESTING

The fact that one is unable to produce enough gravity concentrate at the bench scale for meaningful testing represents a limitation to the hydrometallurgical test program. Because of this, flotation concentrate from the MPTM operation was used in the test program. However, as indicated from the mineralogical analysis, the composition of gravity concentrate is different from flotation concentrate. The objective of the hydrometallurgical testing was to characterize the APT circuit and identify possible areas that may cause grade-recovery limitations in leaching, purification, solvent extraction, and crystallization.

The maximum extraction of tungsten using a caustic leach was reported as 96%, however, the leach conditions were not optimized. A 10 molar caustic bake is used commercially on concentrates (containing 10-40% $WO_3$) based on the stoichiometry of the caustic reaction with wolframite. The main objective of the digestion unit operation is to selectively dissolve tungsten as sodium tungstate.

In order to maintain the solubility of sodium tungstate, a water leach is employed after digestion. The impurities remain virtually unleached in the solid residue but molybdenum extraction was found to be higher than desirable. The weight loss of leached solids was found to be approximately 30%, and the leach liquor contained predominantly iron and manganese hydroxides.

A single stage purification test of the leach solutions indicated that magnesium chloride is effective in removing 65-99% of the arsenic, zinc, and aluminum The levels of lead and phosphate were below the limits of detection in the leach solutions. The removal of molybdenum from the solution was limited to 16%. The use of ammonium hydroxide did not appear to improve the precipitation of heavy metals although ideal conditions for purification were not studied. Purification tests were done at a pH of 10, at 90-90°C for two hours and the loss of tungsten in purification was estimated to be 5.4 to 7.7% of the tungsten contained in the leach feed. Test results suggests that purification may limit tungsten recovery in the APT hydromet.

The solvent extraction testing was limited to shake tests using an organic makeup of 12% alamine 336 (sulfated by sulfuric acid), 12% tributyl phosphate, and 76% kerosene to extract the sodium tungstate from the purified solutions and subsequently produce APT by the reaction of ammonium hydroxide and tungstate in the stripping stage. Preliminary solvent extraction tests done at room temperature and at a pH of 2 suggest that an extraction in excess of 98% could be achieved using two stages derived from a McCabe Thiele diagram. Although additional testing done on a purified solution at a pH of 8 was successful in achieving a good extraction, the recovery of tungsten in stripping was limited. Optimization of process conditions for extraction and hydrolysis reactions in the stripping were not studied.

Analysis of the APT crystals by x-ray diffraction indicated a high purity product. The limited recovery of APT may have been due to an incomplete hydrolysis reaction and/or incomplete crystallization. It is expected that the level of impurities may increase as the product recovery improves.

## 8.0 CONCEPTUAL COMMERCIAL PLANT FLOWSHEET DESIGN AND ECONOMICS

The conceptual feasibility of integrating an APT manufacturing plant with the existing Mount Pleasant Tungsten facility required an investigation of the technical and economic aspects of APT production. The test work discussed in section 7.0 identified the necessary unit operations to produce APT from the low-grade tungsten ores and has indicated some potential areas for further test work. It is also recognized that the definition of various process parameters and a detailed mass balance will be required to establish an order of magnitude estimation of the equipment sizing and cost. This study is intended to identify whether or not further development of this concept is worth pursuing and consequently the areas that require detailed assessments.

The APT production process is based on the integration of grinding, gravity separation, and caustic hydrometallurgy. The process flow for physical separation has been designed to recover the tungsten before liberated grains are over-ground. The primary upgrading may be considered unique although the hydrometallurgical operations to treat the concentrate are standard industrial practice. Since the details of the APT hydromet process design and commercial operating methods have not been published, the preliminary design calculations are intended to provide a basis for a cost analysis and a preliminary technical evaluation.

### 8.1 PROCESS DESCRIPTION

The integrated process is designed to convert crushed ore in the particle size range of -16 mm containing 0.34% $WO_3$ to a primary concentrate containing approximately 16% $WO_3$ and subsequently by chemical methods, produce APT grading in excess of 83.6% $WO_3$. A preliminary design for a commercial operation is based on the throughput of 100 mt/hr of run-of-mine ore. Assuming a plant availability of 85% and an overall recovery of 73.9%, the net production of APT is

2,235,289 kg/yr. The overall process is designed to treat the ore in the most cost effective manner. As illustrated in Drawing No. 0100-01-01 of Appendix D, the proposed plant would include the following unit operations:

- o Primary Screening and Separation.
- o Primary Grind and Separation.
- o Regrind and Separation.
- o Secondary Grind and Separation.
- o Fines Recovery.
- o APT Hydromet.

The following subsections provide a brief description of these circuits, and process diagrams are also included in Appendix D.

### 8.1.1 PRIMARY SCREENING AND SEPARATION

The test program indicated that significant amounts of tungsten can be recovered directly from crushed ore. The oversize material is removed from the crushed ore feed on a continuous basis and the tungsten in the undersize (4 mesh) is separated by a high capacity jig. The operating conditions for the jig are set for optimum grade, and the concentrate is sent to the regrind circuit. The oversize (30 mesh) is removed from the tailings and reground in a secondary rod mill.

### 8.1.2 PRIMARY GRIND AND SEPARATION

The pilot testing done by MPTM using spirals as part of the grinding circuit, provides a design basis for the secondary grind circuit. The ore is ground to approximately minus 14 mesh and is separated after grinding in a closed loop configuration with the jig. The spirals produce a dirty concentrate and the tailings are deslimed and cleaned by reichert cones before secondary grinding.

### 8.1.3 REGRIND AND SEPARATION

The regrind circuit is designed to treat the concentrate recovered from primary circuits. The tonnage is low and the average feed grade is around 8% $WO_3$, containing liberated tungsten and middling. The regrind ball mill is operated in a closed circuit with a DSM screen, treating the undersize (-500 microns) with spirals. A final concentrate is produced by this circuit and the tailings are subsequently treated in the secondary grinding circuit.

### 8.1.4 SECONDARY GRIND, SEPARATION AND FINE RECOVERY

The product from the secondary ball mill is classified by cyclones and the undersize ranging from -100 to -150 microns is classified for coarse, intermediate and fine particle size gravity separation. The tailings-middlings from the rougher of the coarse tables and the middlings from the intermediate tables is reground. The fines (-25 microns) are treated in a separate circuit, and the operating conditions can be selected for an optimum recovery. This conceptual flowsheet design should reduce the amount of tungsten in the fines circuit.

### 8.1.5 APT HYDROMET

A flotation circuit is included in the flowsheet to remove sulfides from the bulk gravity concentrate. The gravity concentrate is reground in a closed circuit using a coarse flotation cell on the cyclone underflow to remove coarse lead and arsenic. An additional bank of flotation cells is included to maximize the recovery and cleaning of bulk sulfides (to release entrained tungsten). It is uncertain if the flotation of sulfides is required, however flotation has been employed by the Chinese in various concentrators for cleaning of gravity concentrates.

The concentrate is thickened and continuously added to a mixer-digester. A caustic solution (50% NaOH) is then added in stoichiometric portions. The total residence time for digestion is four hours at 100°C. A water leach is used after digestion to dissolve the sodium tungstate and to complete the reaction. The discharge from the leach circuit is cooled to 55°C to prevent flashing of the solution in the drum filters.

The leach residue is washed by two stages of filtering with an intermediate slurry repulp. The residue solids are subsequently treated in a tailings treatment circuit and the solution containing sodium tungsten is purified.

In the purification circuit, the impurities (such as silica, arsenic, molybdenum, and phosphorus) are removed by the reaction of magnesium chloride with sodium complexes. Magnesium chloride addition is based on stoichiometry. The following purification conditions have been based on published literature:

$MgCl_2$, 80-90°C, pH 10.5-11.5 (Removal of Si, Fe, Al, Ti)
$MgCl_2$ + $NH_4OH$, 70-80°C, pH 8-9 (Removal of P, As)

The pH of the purification solutions is modified with sulfuric acid. Precipitated solids are removed from sodium tungstate solutions by a continuous pressure filter. Periodically the solids are washed from the filter and transferred to tailings treatment.

The impurity-free tungstate solution is pumped to solvent extraction (SX) and the pH is lowered to 2. Two stages of extraction are required to achieve approximately 98% extraction of the tungsten. An additional stage is used as a quality assurance step.

The loaded organic is stripped with an ammonium hydroxide solution and is washed with deionized water to remove any residual ammonia. Sulfate bonded to the organic is replenished with the addition of sulfuric acid and the by-product sodium sulfate, is treated in the tailings circuit. Strip liquors from the SX circuit

is pumped to the crystallizer to continuously remove APT from the solution. A purge stream from the crystallizer is recycled to purification.

APT crystals are washed and dried at 70-75°C to prevent decomposition. Packaging of the final product is done in a contamination free environment using 150-160 kilogram drums.

## 8.2 PROCESS DESIGN PERSPECTIVE

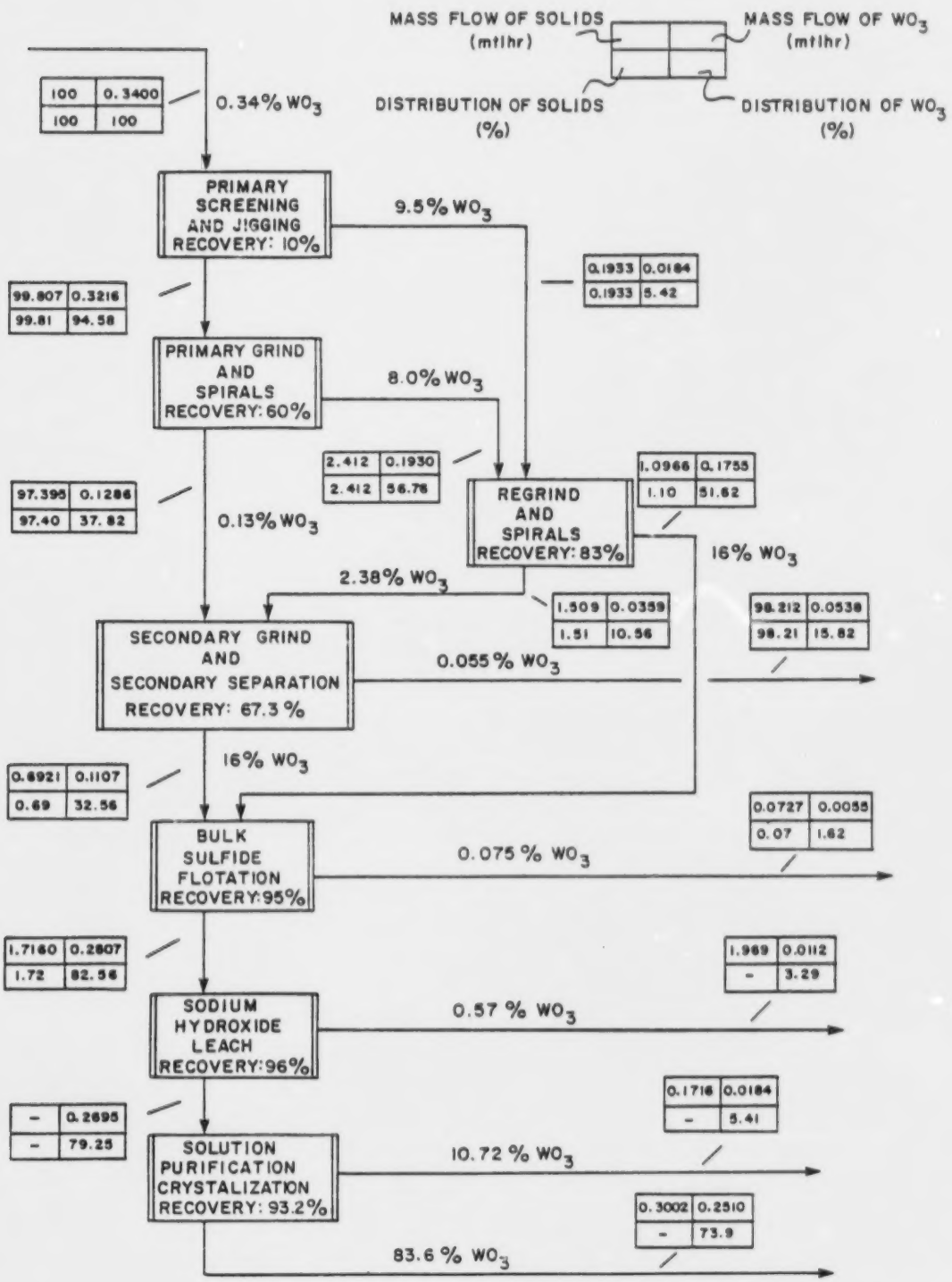
As noted earlier, preliminary design calculations for the production of 2,235 mt/yr of APT from run-of-mine ore have been summarized in Appendix A. The optimum conditions for both gravity separation and the APT hydromet circuits have not necessarily been used in the preliminary design calculations due to the lack of detailed process data. It should be noted, however, that the major uncertainty in the design will be associated with equipment sizing, since a number of assumptions have been made concerning the mass balance and operating conditions.

### 8.2.1 MASS BALANCE

A preliminary mass balance for the entire process has been summarized in Figure 1, giving the approximate distribution of solids and tungsten in major process streams. The assumptions for the mass balance are summarized in Appendix A and are based on the bench scale testing, pilot testing, and on standard industrial operating practice.

The main objective of integrating grinding with gravity separation was to improve the overall recovery of primary upgrading. The recovery of primary upgrading has been estimated to be 82.6% for a grade of 16% $WO_3$. It is believed that the major loss of tungsten will be in the fine recovery circuit, and that the recovery

# MASS BALANCE SUMMARY



of tungsten in the secondary grind circuit will be in the range of 64.5-70.2% depending of the amount of slime tungsten produced. The distribution of slime tungsten (-25 micron) may be reduced to 25-35%, which is a major improvement over single stage grinding.

The predicted mass balance for a commercial hydromet circuit has been based on bench scale tests. The major loss of tungsten is expected to be associated with leaching and solution purification unit operations. The extraction and recovery of tungsten in the leaching circuit may be improved, however, high losses of tungsten in commercial purification operations have been reported in the literature. The projected recovery of tungsten in the hydromet, calculated to be 89.5%, is low by industrial standards. Thus, the calculated plant recovery of 73.9% may be conservative.

### 8.2.2 EQUIPMENT SIZING

The sizing calculations have been simplified for an order of magnitude estimate. Although design estimates are included Appendix A, they do not provide a basis for commercial design. The assumptions for equipment sizing and has been summarized in Appendix A and the equipment requirements are summarized in Table 1.

The equipment requirements were based on an assumed integration with the existing processing facility at Mount Pleasant. The equipment list has taken into account the requirements for grinding, gravity and APT hydromet circuits. It was assumed that additional processing equipment and buildings would not be required to commercialize this process at Mount Pleasant. Additional requirements for a complete APT processing facility would include:

- Utilities (compressed air, steam, water treatment).
- Crushing and ore storage.
- Off-gas treatment (ammonia scrubbing).
- Effluent and solid tailings treatment.
- Tailings pond.
- Product drying and packaging.

Table 1: Mechanical Equipment List Page 1

| Equipment Number | Mechanical Equipment | No. Each | Size | Rating | H.P./RPM EA. | Material |
|---|---|---|---|---|---|---|
| Area 2100 | Primary Screen (-4 mesh) | 1 | 4' x 8' | 40 mt/hr | 5.0 Hp | CSRL |
| | Secondary Screen (-20 mesh) | 1 | 6' x 12' | 5 mt/hr | 10.0 Hp | CSRL |
| | Primary Rod Mill | 1 | 10.5' x 14.5' | 110 mt/hr | 700 Hp | CSRL |
| | Tertiary DSM Screen (14 mesh) | 4 | 4' x 6' | 100 mt/hr | | CS |
| | Regrind DSM Screen (32 mesh) | 1 | 4' x 6' | 5-10 mt/hr | | CS |
| | Primary Jig and Distribution Box | 4 | 24" x 36" | 20 mt/hr | 2.0 Hp | CSRL |
| | Primary Spirals and Distribution Box | 63 | | 2 mt/hr | | URF |
| | Regrind Spirals and Distribution Box | 5 | | 2 mt/hr | | URF |
| | Regrind Ball Mill | 1 | 5' x 5' | 5 mt/hr | 35.0 Hp | CSRL |
| | Coarse Concentrate Pump, Sala Vertical (SV) | 1 | | 2 m3/hr | 1.0 Hp | CSRL |
| | Jig Tailing Pump and Pump Box | 1 | 3" x 3" | 70 m3/hr | 15.0 Hp | CSRL |
| | Rod Mill Discharge Pump and Pump Box | 1 | 4" x 5" | 150 m3/hr | 35.0 Hp | CSRL |
| | Primary Spiral Feed Pump and Pump Box | 2 | 4" x 5" | 150 m3/hr | 35.0 Hp | CSRL |
| | Regrind Feed Pump (SV) | 1 | | 2 m3/hr | 1.0 Hp | CSRL |
| | Regrind Concentrate Pump (SV) | 1 | | 2 m3/hr | 1.0 Hp | CSRL |

Table 1: Mechanical Equipment List Page 2

| Equipment Number | Mechanical Equipment | No. Each | Size | Rating | H.P./RPM EA. | Material |
|---|---|---|---|---|---|---|
| Area 3100 | Primary Cyclones and Distributor | 1 | 26" | 272 m3/hr (5 psi) | - | CSRL |
| | Secondary Cyclones and Distributor | 6 | 15" | 90 m3/hr (10 psi) | - | CSRL |
| | Tertiary Cyclones and Distributor | 96 | 4" | 8.0 m3/hr (15 psi) | - | CSRL |
| | Secondary Ball Mill | 1 | 11.5' x 11.0' | 98 mt/hr | 750 Hp | CSRL |
| | Reichert Cones and Distribution Box | 3 | 50 mt/hr | 79 mt/hr | - | URP |
| | Coarse Separation Tables (Double Deck) & Dist. Box | 53 | 2 mt/unit | 90 mt/hr | 2 x 1 Hp | PE/CS |
| | Intermediate Separation Tables (Double Deck) & Dist. Box | 44 | 2 mt/unit | 83 mt/hr | 2 x 1 Hp | PE/CS |
| | Fine Separation Tables (Double Deck) & Dist. Box | 11 | 2 mt/unit | 15 mt/hr | 2 x 1 Hp | CSRL |
| | Primary Cyclone Feed Pump & Pump Box | 1 | 5" x 4" | 120 m3/hr | 30 Hp | CSRL |
| | Secondary Ball Mill Feed Pump & Pump Box | 1 | 8" x 6" | 225 m3/hr | 50 Hp | CSRL |
| | Secondary Cyclone Feed Pump & Pump Box | 2 | 8" x 6" | 275 m3/hr | 50 Hp | CSRL |
| | Secondary Concentrate Transfer Pump (SV) | 1 | | 2 m3/hr | 1.0 Hp | CSRL |
| | Tertiary Cyclone Feed Pump & Pump Box | 2 | 6" x 5" | 200 m3/hr | 50 Hp | CSRL |
| | Fine Tailing Recycle Pump & Pump Box | 1 | 1 1/2" x 1 1/2" | 10 m3/hr | 2 Hp | CSRL |
| | Tailing Transfer Pump & Pump Box | 4 | 8" x 6" | 225 m3/hr | 50 Hp | CSRL |

Table 1: Mechanical Equipment List Page 3

| Equipment Number | Mechanical Equipment | No. Each | Size | Rating | H.P./RPM EA. | Material |
|---|---|---|---|---|---|---|
| Area 4100 | Gravity and Tungsten Concentrate Thickener | 2 | 2.0 m dia | 3.4 m3/hr | 5.0 Hp | CSRL |
| | Process Water Tank | 1 | 2.4 m dia x 2.0 m | 12 m3 | | |
| | Gravity and Tungsten Concentrate Holding Tank & Agitator | 2 | 1.6 m dia x 1.9 m | 3.7 m3 | 3.0 Hp | CSRL |
| | Conditioning Tank and Agitator | 1 | 0.8 m dia x 1.0 m | 0.5 m3 | 0.5 Hp | CSRL |
| | Coarse Flotation Cell | 1 | 2 m3 | 1 mt/hr | 1.0 Hp | CSRL |
| | Regrind Cyclone | 1 | 3" dia | 1.5 m3/hr | | CSRL |
| | Regrind Mill | 1 | 3' x 3' | 1 mt/hr | 10.0 Hp | CSRL |
| | Tungsten Concentrate Belt Filter | 1 | 3.5 m2 | 0.69 mt/hr | 4.0 Hp | CS |
| | Caustic Digestor | 1 | 1.6 m3 | 0.3363 m3/hr | 65.0 Hp | 316SS |
| | Water Leach Tanks, Agitators and Steam Coils | 3 | 0.76 dia x 0.92 m | 0.40 m3 | 1.0 Hp | FRP |
| | Leach Residue Heat Exchanger | 1 | | | | CS |
| | Leach Residue and Repulp Tanks | 3 | 1.3 dia x 1.6 m | 2.1 m3 | 1.5 Hp | CSRL |
| | Leach Residue Drum Filters & Vacuum System | 2 | 5 m2 | 0.5 mt/hr | 3.0 Hp | CS |
| | Thickener Underflow Pumps (ODS) | 2 | 2" | 1.6 m3/hr | | CSRL |
| | Process Water Pump (Hydromet) (with spare) | 2 | 1 1/2" x 1 1/2" | 12 m3/hr | 2.0 Hp | CS |
| | B/S Regrind Transfer Pumps (SX) | 2 | | 2 m3/hr | 1.0 Hp | CSRL |
| | Gravity and Tungsten Conc. Transfer Pumps | 2 | 1" x 1" | 1 m3/hr | 0.5 Hp | CSRL |
| | Leach Residue Transfer Pump and Repulp Pumps | 3 | 1" x 1" | 1 m3/hr | 0.5 Hp | CSRL |
| | Filtrate Pumps | 2 | 1" x 1" | 0.5 m3/hr | 0.25 Hp | PP |

Table 1: Mechanical Equipment List Page 4

| Equipment Number | Mechanical Equipment | No. Each | Size | Rating | H.P./RPM EA. | Material |
|---|---|---|---|---|---|---|
| Area 4100 | Purification Reagent Storage Bins | 2 | 3.8 dia x 7.9 m | 85 m3 | | CS |
| | Solution Purification Tanks and Agitators | 6 | 0.7 dia x 0.84 m | 0.3 m3 | 0.5 Hp | FRP |
| | Purification Holding Tank and Agitators | 2 | 1.3 dia x 1.6 m | 2.04 m3 | 3/4 Hp | FRP |
| | Precipitate Pressure Filters | 2 | | | | FRP |
| | Sodium Tungstate Holding Tank and Agitator | 1 | 1.6 dia x 1.9 m | 3.68 m3 | 1.25 Hp | FRP |
| | Solvent Extraction Cells, Agitators and Pumps | 5 | | 0.28 m3/0.14 m3 | 0.25 Hp | FRP |
| | Organic Sulfation Tank and Agitator | 1 | 0.7 dia x 0.8 m | 0.25 m3 | 0.25 Hp | FRP |
| | Wash Raffinate Holding Tank | 1 | 1.1 dia x 1.4 m | 1.37 m3 | | FRP |
| | Crystallizer, Heat Exchanger, and Condenser | 1 | 1.4 m3 | 207 kg/hr | 1.0 Hp | 316LSS |
| | Product Drum Filter | 1 | 6.2 m2 | 306 kg/hr | 1.5 Hp | 316SS |
| | Purification Reagent Feeder | 2 | | 72 kg/hr | 0.5 Hp | CS |
| | Precipate Filter, SX Feed, and Crystallizer Feed Pump | 4 | 1" x 1" | 0.407 m3/hr | 0.5 Hp | 316SS |
| | Filtrate Pump (Product Dewatering) | 1 | 1/2" x 1/2" | 0.03 m3/hr | 0.25 Hp | 316SS |

## 8.3 PROCESS ECONOMICS

An order of magnitude capital and operating cost estimate was determined from the preliminary conceptual design. The cost estimates have been prepared for the purpose of evaluating the economic feasibility of APT production and to identify the areas having economic limitations.

### 8.3.1 CAPITAL COST ESTIMATE

Preliminary capital cost estimates have been derived from budget pricing received from various suppliers over the last three years. The budget prices have been factored according to the equipment cost index M&S, published by Chemical Engineering Magazine. The applicability and cost of major capital items such as grinding mills, gravity separation equipment, digesters, solvent extraction cells, and crystallizers have been discussed with equipment suppliers. The cost for the remainder of the plant equipment has been derived by applying generally accepted industry estimating guidelines, adjusted for projected size and duty.

A capital cost estimate has been summarized in Appendix B. Transportation and installation costs have been factored in most cases. Electrical, instrumentation, and piping for construction have been factored as a percentage of the purchased equipment cost. Details of the cost breakdown and assumptions have been included in Appendix B.

The overall capital cost for procuring and constructing grinding/gravity/APT hydromet circuits at the Mount Pleasant facility has been estimated at approximately $28 million (January, 1989, Canadian funds) with a level of confidence of +40% -5%. Table 2 provides a breakdown of the major components in this capital cost. The total capital investment does not include:

- Equipment for utilities, tailings treatment, crushing, drying, and packaging process.

**TABLE 2**

**TOTAL CAPITAL INVESTMENT PRELIMINARY ESTIMATE**
**2,400 MT/DAY TUNGSTEN ORE**
**MOUNT PLEASANT, NEW BRUNSWICK**

**GRINDING/GRAVITY/APT HYDROMET**

**January, 1989**

| | |
|---|---|
| Purchase of Equipment | 10,238,000 |
| Transportation and Installation Cost | 3,531,900 |
| Electrical | 1,431,000 |
| Instrumentation | 1,300,000 |
| Piping | 2,080,000 |
| **TOTAL DIRECT COST** | **18,580,900** |
| Engineering Supervision and Construction Mgt. (20%) | 3,716,000 |
| Construction Indirects (3%) | 557,000 |
| Startup Service and Vendor Supervisor (4%) | 743,000 |
| Contingency (10%) | 1,858,000 |
| **TOTAL INDIRECT COST** | **6,874,000** |
| Total Fixed Cost | 25,454,900 |
| Working Capital (10% of Total Fixed Cost) | 2,545,000 |
| **TOTAL FIXED CAPITAL INVESTMENT** | **27,999,900** |

LEVEL OF CONFIDENCE: +40%, -5%
CHEMICAL ENGINEERING PLANT COST INDEX: 340

( ) Percentage of Total Direct Cost

NOTE: Costing is based on an integration with the existing Mount Pleasant facility.

- o Building and storage facilities.
- o Mining related capital expenditures.
- o Infrastructure for transportation, waste disposal, and other non-process related items.
- o Electricity generation and distribution outside the plant.
- o Acquisition of the site and site improvements.
- o Process development costs (either licensing or research and development expenses).

The level of confidence should be considered preliminary because of the lack of an accurate mass balance and availability of details of process operating variables. The equipment cost of $10.2 million is predominantly associated with the cost of the grinding and gravity separation circuits. The APT hydromet accounts for approximately 14% of the total equipment purchase cost, or approximately $1.4 million.

For the same scale of operation, the proposed cost for an APT hydromet may be as high as $3.9 million. The major reason being, the recycle of solutions in the hydromet and the use of a significant amount of wash water may increase the size of the hydromet by 2-3 fold. If this is the case, the total fixed capital investment may be as high as $34 million. Additional errors in the costing of the gravity circuit would be related to the change in the grind characteristics, causing the mass balance to shift, thus changing the equipment sizing requirements for each circuit.

### 8.3.2 OPERATING AND MAINTENANCE COSTS

An estimate of production costs, basis for estimates and assumptions are summarized in Appendix C. Exclusive of capital, a series of key assumptions have been made:

- o Operation of the entire plant (excluding crushing and packaging) would be on a 7 day per week, 24 hour basis.
- o Operating costs for crushing, tailings treatment, product drying, and packaging are based on factored costing from similar commercial operations.
- o Staffing includes 71 employees for the processing plant including management, operators, and technical and maintenance personnel with an average wage of approximately $33,000 per annum. An allowance has been made for overtime (10%) and burdens (32%) in addition to annual salaries and payroll.
- o Operating and maintenance supplies are based on estimated requirements for standard industrial practice for foreseeable items. Reagent costs are based on market values F.O.B. Montreal.
- o The operating costs do not include a value for the run-of-mine ore or administration.

The total direct operating cost, exclusive of capital charges, taxation, and royalties, has been estimated to be $13.497 per tonne of ore processed or $4.50 per kilogram of APT produced, based on a plant operating time availability of 85%. A breakdown of the projected operating costs are shown in Table 3 with a level of confidence of +30% -5%.

A major cost component for APT production is operating and maintenance labour expenses, estimated at 32.7% of the total operating costs. The complexity of the flowsheet in the gravity and hydromet circuits may require additional operators for the duration of commissioning. However it is anticipated that the use of automation with floor and control room operators will provide satisfactory control of the plant. In anticipation of operating difficulties, the total expenditures for manpower could be increased by 10%. This contingency was not included in operating cost estimates.

Reagents, operating and maintenance supplies, and crushing and tailings treatment each contribute 10-15% to the total operating cost. It is not expected that operating and maintenance costs would increase based on a detailed design. Reagent consumption may increase in purification and hydrolysis, depending on the nature of the commercial operation. If the reagent consumption was to double (because of the projected increase in solution volume) for these unit operations,

**TABLE 3**

**OPERATING COST PRELIMINARY ESTIMATE**
**2,400 MT/DAY TUNGSTEN ORE**
**MOUNT PLEASANT, NEW BRUNSWICK**

**GRINDING/GRAVITY/APT HYDROMET**

**January, 1989**

| | Annual Basis | Cost Per Metric Tonne of Ore (744,600 MT/YR)* | Distribution of Cost (%) |
|---|---|---|---|
| Labour (71 employees) | 3,287,000 | | 32.7 |
| Reagents | 1,358,300 | | 13.5 |
| Operating & Maintenance Supplies | 1,250,000 | | 12.4 |
| Electrical Power | 473,600 | | 4.7 |
| Water and Steam | 99,700 | | 1.0 |
| Crushing | 1,041,700 | | 10.4 |
| Product Packaging | 271,300 | | 2.7 |
| Tailing Treatment | 1,182,600 | | 11.8 |
| Contractors | 320,000 | | 3.2 |
| Insurance on Equipment | 122,500 | | 1.2 |
| Transportation/R&D/Marketing | 643,000 | | 6.4 |
| **TOTAL** | **10,049,700** | **13.497** | **100.0** |

LEVEL OF CONFIDENCE: +30%, -5%
CHEMICAL ENGINEERING PLANT COST INDEX: 340

* NOTE: BASED ON AN 85% AVAILABILITY, THE PRODUCTION OF APT IS APPROXIMATELY 2,235,289 KG/YR

the total cost for reagents would increase by 1.5 fold, and the net increase in the operating cost would be 10%. Therefore, the level of confidence of +30% is considered realistic for a preliminary costing.

## 8.4 DEVELOPMENT OPTIONS FOR TUNGSTEN

The preliminary conceptual design and cost estimates for an APT manufacturing plant has provided data for an order of magnitude assessment of the process economics. The process economics is based on integrating the proposed processing concept with a mining, ore storage, crushing, product packaging and utilities at the existing Mount Pleasant facility. If the net value of the ore is estimated at $16 per metric tonne (including mining and administration cost) and assuming a return on investment of 15% (before taxes based on ten years), the value of the APT would have to be in excess of $CDN 150/MTU $WO_3$. The present value of APT is $75 to $85/MTU $WO_3$ and is sometimes sold at even lower prices by Chinese producers.

Recent studies by Hard Metal Research Inc. of Mississauga, Ontario, reviewed the technical and economic feasibility of producing tungsten powder, tungsten carbide and hard metal products from APT. The total capital investment for the treatment of 1,350 metric tonne per year was estimated between $25 and $50 million (1987 dollars) for a reduction, carbonization and sintering plant. These additional capital expenditures would increase the value of the final product by about five fold and the return on investment after taxes, would be between 9 and 19%. The complete integration of a processing facility for the production of tertiary tungsten products has been identified as a viable investment opportunity, moreso than the sole production of APT.

## 9.0 CONCLUSIONS

The data obtained from this preliminary study does not support the production of APT at present day pricing. It is evident that a fully integrated processing facility to produce tungsten carbide at the mine site may be economically viable, where the production of APT from ore would be an intermediate processing step. The technical aspect of such a facility would require a detailed assessment.

The economic feasibility of processing tungsten ores in New Brunswick and elsewhere in North America, can be strongly altered by the actions of producers in China. A detailed understanding of the strengths and weaknesses of the Chinese industry would be an important prerequisite to developing an integrated processing facility.

In summary, is evident that additional development costs would be associated with the determination of a definite design and an overall feasibility of APT production. The continued development of the Mount Pleasant ores may be justified by the production of tungsten carbide and/or hard metals but further studies would be required to determine the overall feasibility of an integrated complex. The major technical limitations in the development of Mount Pleasant ores are the loss of fine tungsten and the removal of impurities. The production of APT from ores as an intermediate step for tungsten carbide production has a potential to overcome these limitations. Tungsten carbide and powder may be a viable commercial development concept for Mount Pleasant if high purity APT can be manufactured at, or near, the mine site.

## 10.0 RECOMMENDATIONS

Based on the current conceptual design, the next step in the development of an APT processing circuit would be to further define the process parameters and complete a feasibility study for a commercial plant. Some unique features of a gravity/APT hydromet have been identified, however it is recommended that the following critical aspects be part of technical development plans for Mount Pleasant ores:

(i) Further investigations into chlorination technology should be established to determine the competitiveness of this technology with conventional APT processing methods.

(ii) There is no distinction between the grinding and gravity separation in the proposed flowsheet. The net effectiveness of this method of treating the ores should be determined to define the net reduction in fines loss.

(iii) A complete assessment of process flow configuration and equipment requirements for gravity separation will be required (i.e., a rougher-cleaner may be required for jig separation of crushed ore).

(iv) The economic grade for gravity concentrate should be determined based on the size of a hydromet circuit, the increased recovery of gravity separation and the net change in operating costs.

(v) The use of pressure leach on low-grade concentrates has increased the interest in integration of tungsten processing from ore to carbide. Various routes to selectively leach tungsten and reduce impurity levels should be investigated.

(vi) The use of ore sorters for primary upgrading of low ore grades should be investigated. Recent improvements in x-ray sensors have been noted in the literature.

(vii) Further research is required to improve the efficiency of purification of sodium tungstate solutions and to optimize tungsten recovery in purification. For example, phosphorous and silica can be removed by magnesium and aluminum sulfates, arsenic and molybdenum can be precipitated as a sulfide by the addition of sodium sulfide. Ideal conditions, reagents, number of stages, and process flow configuration need to be defined.

(viii) Solvent extraction is normally run at a pH of 2 for optimum extraction and at 50°C to prevent silica precipitation. Ideal

operating conditions and reagents used for solvent extraction should be tested, basing the evaluation on rejection of impurities, cost of reagents, and stability of operation.

(ix) The need for silica precipitation on strip solutions before crystallization will have to be determined based on product quality. Conditions for silica precipitation are dependent on ammonia concentrations, time, temperature, pH, and reagents used.

(x) In order to improve on tungsten recoveries in the SX strip solutions, an excess of ammonium will be required. The ammonium balance will have to be designed based on recycle from crystallization. The need for ammonia strip, before crystallization will have to be identified.

(xi) The recycle of tungstate solutions, the use of wash solutions, and the optimization of the crystallizer are all dependant on water balance. The ideal operating balance for APT hydromet will be required for precise equipment sizing.

(xii) The limitations and cost of mining the ore and the variability in quality need to be identified for a complete assessment of the project.

(xiii) Effluent treatment from leach residue, purification precipitates and ammonia scrubbing should be assessed.

The development of a fully integrated processing facility will require a large amount of capital. The feasibility of such a venture should be considered based on the return on investment, existing infrastructure, security of feedstock, and market potential. It is recommended that all of these factors be included in further investigations to establish a development plan for Mount Pleasant.

**11.0 BIBLIOGRAPHY**

ABEIDU, A. M. Flotation Studies on Wolframite. IMM Transactions, March 1975, Section C, Vol. 84, No. 820.

BETZ, L. D. et al. "Removal of Silica from Water by Cold Process." Industrial and Engineering Chemistry, October, 1940, Vol. 32, No. 10, pp 1320-1323.

----------. "Removal of Silica from Water by Hot Process." Industrial and Engineering Chemistry, October, 1940, Vol. 32, No. 10, pp. 1323-1329.

BLIZZARD, B. Ball Mill Grind and H.G.M.S. on Gravity Concentrate. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, July 21, 1987.

----------. Gravity and H.G.M.S. on Mount Pleasant Ores. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, April 9, 1987.

----------. H.G.M.S. on Mount Pleasant Ores. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, April 20, 1987.

----------. Identification of Rare Earths and Tungsten Loss with Gravity Separation. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, July 28, 1987.

----------. Phosphate Float on Ore. Process Studies Report, New Brunswick Department of Natural Resources and Energy, New Brunswick, June 15, 1987.

BURROWS, J. C. "Tungsten Supply and Demand Examined." Mining Journal, April 4, 1986, pp. 248-249.

CASSELLA, VINCENT and IRANI, MAZIN R. Removing Metal Ions from Aqueous Effluents. United States Patent, No. 4,200,528, April 29, 1980.

CHIN, P. C. et al. A New Slime Concentrator -- The Rocking-Shaking Vanner. Developments in Mineral Processing, 2: Part B, Elsevier Scientific Publishing Company and PWN-Polish Scientific Publishers, 1981, p. 1398-1423.

COETZEE, C. F. B. and FIEBERG, M. M. Hydrochloric Acid Leaching and Low-Temperature Roasting of Tin-Tungsten Concentrates from Van Roois Vley. MINTEK Report No. M93D, Council for Mineral Technology, April 22, 1983.

DAVISON, J. G. Mineralogical Studies of Mount Pleasant Tungsten Ore. Witteck Project No. 5295 for the Department of Natural Resources and Energy, New Brunswick, Mineral Resources Division. 1987.

DROBNICK, J. L. and LEWIS, C. J. "Recovery and Purification of Tungsten by the Liquid Ion Exchange." Unit Processes in Hydrometallurgy, 1963, pp. 504-515.

HARRIS, R. D. Pre-Feasibility (Economic) Review of Mt. Pleasant Fire Tower Zone Deposit. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, April 15, 1987.

HO, C. ERIC. "Tungsten." Mining Magazine, March, 1987, pp. 219-221.

HOGSETT, ROBERT F. et al. Process for Conditioning Tungsten Concentrates. United States Patent, No. 4,338,287, July 6, 1982.

MACINNIS, M. B. and KIM, T. K. "Commercial Processes for Tungsten and Molybdenum." Handbook of Solvent Extraction, Lo, Baird, and Hanson, 1983, pp. 689-694.

MERCADE, VEN. "New Pathways to Tungsten Flotation." E&MJ, August, 1983, pp. 52-58.

O'DONNELL, MITCH. Mineralogical Identfication of a Final Product from the Ammonium Paratungstate Flowsheet. Technical Notes, Department of Natural Resources and Energy, New Brunswick, July 11, 1988

----------. Processing of Specialty Metals in New Brunswick. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, May 12, 1988.

OTTLEY, DEREK J. "Technical, Economic and Other Factors in the Gravity Concentration of Tin, Tungsten, Columbium, and Tantalum Ores." Minerals Sci. Engng., April, 1979, Vol. 11, No. 2, pp. 99-121.

QUENEAU, PAUL B. and BERTHOLD, CORNELIUS E. "Silica in Hydrometallurgy: An Overview." Canadian Metallurgical Quarterly, 1986, Vol. 25, No. 3, pp. 201-209.

Quatrimi, Lucretia R. et al. Tungsten Recovery from Tungsten Ore Concentrates by Caustic Digestion. United States Patent, No. 4,353,878, October 12, 1982.

RITCEY, G. M. and ASHBROOK, A. W. "Tungsten." Solvent Extraction Principles and Applications to Process Metallurgy: Part III, pp. 447-451.

SHANLUN, SUN, et al. New Techniques in Processing Tungsten Ore Slimes. Papers Presented at the International Conference "Mineral Processing and Extractive Metallurgy" by the Institution of Mining and Metallurgy and the Chinese Society of Metals, Kunming, Yunnan Province, People's Republic of China, October 27 to November 3, 1984, pp. 521-530.

SPROUL, G. H. Leach Tests on Sb and W Concentrates. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, August 31, 1987.

----------. Review Leach Methods to Extract REO's from MPT Flotation Con-

centrates. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, July 6, 1987.

----------. Separation of Tungsten from Caustic Leach Solution by Solvent Extraction. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, September 1, 1987.

TAYLOR, D. Comparison of Mozley Separator and Other Separators. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, September 3, 1987.

----------. Gravity Separation on Jig, Spiral and Shaking Table. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, September 3, 1987.

----------. Preliminary Design of Gravity Plants. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, September 14, 1987.

----------. To Determine the Optimum Conditions for Gravity Separation. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, August 26, 1987.

TAYLOR, DON and BLIZZARD, BRENT. Grinding Parameters. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, May 30, 1987.

TAYLOR, D. and THIBAULT, D. Grinding Characteristics of Ore. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, June 15, 1987.

THIBAULT, D. Mineralogical Studies on Mount Pleasant Tungsten Ore. Process Studies Report, Department of Natural Resources and Energy, New Brunswick, November 30, 1987.

Tungsten: 1982. Proceedings of the Second International Tungsten Symposium, San Francisco, June 1-5, 1982, pp. 28-80.

"U.K. Tungsten Mine with an Interesting History is Again Producing Concentrates." Mining Magazine, March, 1987, Vol. 136, No. 3, pp. 169-175.

WYLLIE, ROBERT J.M. and PAZOUR, DONALD. "Xihuashan Mines, Mills 3,000 Daily Tons Tungsten Ore."

ZELIKMAN, A. N. "Magnesium, Titanium, Rare Metals, Semiconductors." Tsvetnye Metally/Non Ferrous Metals, pp. 60-65.

# APPENDIX A

## PREFEASIBILITY OF AMMONIUM PARATUNGSTATE PRODUCTION AT THE MOUNT PLEASANT MINE

## ORDER OF MAGNITUDE EQUIPMENT SIZING FOR 2,400 MT/D ORE GRAVITY AND AMMONIUM PARATUNGSTATE HYDROMET

## 1.0 BASIS FOR EQUIPMENT SIZING

Production Capacity: 2,400 mt/d
Grade of Ore: 0.34% $WO_3$ (0.27% W)
Fine Ore Crushed to: -16 mm
Recovery of Tungsten in a Primary Gravity Circuit: 82.6%
Overall Recovery of Tungsten for APT Production: 73.9%

## 2.0 SIMPLIFIED MASS BALANCE

### 2.1 PRIMARY SCREEN AND JIGGING

Assumption 1:

Fine ore crushed to -16 mm has about 48% passing 4 mesh based on average crushing circuit analysis (Norberg Reference Manual, page 5-6).

Amount of ore to primary jig = 100 mt/hr x 0.48
= 48 mt/hr

Assumption 2:

Based on coarse ore testing using a jig, a concentrate of 9.5% $WO_3$ was produced at 10% recovery.

Minimum amount of concentrate to regrind circuit
= 48 mt/hr x 0.0034 $WO_3$ x 0.10/(0.095)
= 0.1718 mt/hr

Maximum amount based on a 25% recirculating load = 0.2147 mt/hr

Average flow to regrind = 0.1933 mt/hr solids
= 0.0184 mt/hr $WO_3$

### 2.2 PRIMARY GRIND AND SPIRAL

Assumption 3:

The spiral operation is similar to the test work done at the Mount Pleasant Mine on undersize from the primary DSM screens having 80% passing 900 micron. At a grade of 8.0% $WO_3$, the recovery was reported by A. Boire and L. Duval to be about 60%, (Confidential report on Spiral Concentrators, MPT, November, 1984).

Amount of concentrate to regrind = ((100 mt/hr x 0.0034 $WO_3$/hr)
- (0.0183 mt $WO_3$/hr))
x 0.60/0.080 $WO_3$/mt conc

= 2.412 mt/hr solids
= 0.1930 mt/hr $WO_3$

## 2.3 REGRIND CIRCUIT AND SPIRALS

Total solid feed to grind = 0.1933 + 2.412 mt/hr
= 2.6053 mt/hr solids

Total tungsten to feed to regrind = 0.0184 + 0.1930 mt/hr
0.2114 mt/hr

Grade of 8.11% $WO_3$ with a distribution of 62.18% (plant feed).

Assumption 4:

For the production of concentrate at 16-20% $WO_3$, it is estimated that a recovery of 80-85% may be practical (average recovery 83%).

Amount of concentrate to bulk sulfide flotation
= (0.2114 mt/hr $WO_3$ x 0.83)/0.16
= 1.0966 mt/hr solids

Overall recovery of tungsten based on primary and regrind circuit is
= (1.0966 x 0.16)/0.34 x 100%
= 51.60%

## 2.4 SECONDARY GRIND AND GRAVITY SEPARATION CIRCUIT

Tailing from primary grind circuit = 99.807 mt/hr - 2.412 mt/hr
= 97.395 mt/hr solids

Tungsten in primary grind tails = 0.3216 mt/hr - 0.1930 mt/hr
= 0.1286 mt/hr $WO_3$

Grade of primary grind tails = 0.13% $WO_3$

Tailing from regrind circuit = 2.6053 mt/hr - 1.0966 mt/hr
= 1.5087 mt/hr solids

Tungsten in regrind circuit tails = 0.2114 mt/hr - 0.1755 mt/hr
= 0.0359 mt/hr $WO_3$

Grade of regrind tails = 2.38% $WO_3$

Total feed to secondary grind circuit = 97.395 mt/hr + 1.5087 mt/hr
= 98.9037 mt/hr solids

Tungsten feed to secondary grind circuit = 0.1286 + 0.0359
= 0.1645

Grade of secondary grind feed = 0.17% $WO_3$

Assumption 5:

Based on the test work with the Mosley Mineral Separator, the following information may be used to estimate the recovery of the secondary grind circuit:

a. for material +75 um, a grade of 16-20% $WO_3$ can be achieved having 75-80% recovery (Graph 2, DNR Process Studies Report, No. F-16, September, 1987).

b. for material -75 um, a grade of 15.6% $WO_3$ is achieved with a 58% recovery (Figure 1, DNR Process Studies Report, No. F-15, September, 1987).

NOTE: Limited recovery of -75 um material is due to a large amount of -25 um material containing about 42-63% $WO_3$ and a recovery of 28.8% (Table 4, DNR Process Studies Report, No. F-7, June, 1987 and F-12, July, 1987).

c. for a grind of 100% -150 micron, the distribution of tungsten for -75 um is 78% (Table 4, DNR Process Studies Report, No. F-7, June, 1987).

Case 1:

The secondary grind and separation circuit is designed to reduce the amount of slime (-25 um) tungsten. It is expected that the slime content will be reduced to about 25% and the amount of tungsten passing 75 um will be 45.5% (based on particle size distribution of Table 2, DNR Process Studies Report, No. F-7, June, 1987).

Amount of +75 um tungsten in secondary grind circuit
= 98.9037 mt/hr x 0.0017 $WO_3$ x (1 - 0.455)
= 0.0916 mt/hr $WO_3$

Amount of +75 um tungsten recovered = 0.0916 mt/hr x 0.7750
= 0.0710 mt/hr $WO_3$

Amount of -75 um tungsten recovered
= 98.9037 mt/hr x 0.0017 x 0.445 x 0.58
= 0.0444 mt/hr $WO_3$

Total tungsten recovered in secondary grind = 0.1154 mt/hr $WO_3$

For a grade of 16.0% $WO_3$, the amount of concentrate
= 0.7211 mt/hr solids

Tungsten recovery for secondary circuit = 0.1154/0.1645
= 70.15

Total feed rate at a grade of 16% $WO_3$ to bulk sulfide flotation
= 0.7211 mt/hr + 1.0966 mt/hr
= 1.8177 mt/hr solids

Total tungsten to bulk sulfide flotation = 0.2909 mt/hr $WO_3$

For primary upgrading, the overall recovery = 85.56%

Case 2:

For the slime content ($WO_3$ -25 um) reduced to 35% and the amount of tungsten passing 75 um, will be 74% (based on particle size distribution of Table 3, DNR Process Studies Report, No. F-7, June, 1987).

Amount of +75 um tungsten in secondary grind circuit
= 98.9037 mt/hr x 0.0017 $WO_3$ x (1 - 0.74)
= 0.0437 mt/hr $WO_3$

Amount of +75 um tungsten recovered = 0.0437 mt/hr x 0.7750
= 0.0339 mt/hr $WO_3$

Amount of -75 um tungsten recovered
= 98.9037 x 0.0017 x 0.74 x 0.58
= 0.0722 mt/hr $WO_3$

Total tungsten recovered in secondary grind = 0.1061

For a grade of 16.0% $WO_3$, the amount of concentrate
= 0.6631 mt/hr solids

Tungsten recovery for secondary circuit = 64.50%

Total feed rate at a grade of 16% $WO_3$ to bulk sulfide flotation
= 0.6631 mt/hr + 1.0966 mt/hr
= 1.7597 mt/hr solids

Total tungsten to bulk sulfide flotation = 0.2816 mt/hr $WO_3$

For primary upgrading, the overall recovery = 82.82%

## 2.5 BULK SULFIDE FLOTATION

Assumption 6:

Tungsten recovery in bulk sulfide flotation is 9, based on commercial operating practice of the Billiton flowsheet.

Average feedrate to bulk sulfide circuit = 1.7887 mt/hr solids

Overall recovery on upgrading = (1.7887 x 0.16 x 0.95)/0.34
= 79.97% recovery of $WO_3$

Based on arsenopyrite, galena and wolframite having a similar specific gravity (in the range of 6.0-7.5), the composition of the gravity concentrate is approximately:

| | Ratio to Wolframite |
|---|---|
| 15.7% $FeAs_2$/FeAsS | 1.06 |
| 14.8% $FeWO_4$ | 1.00 |
| 8.9% PbS | 0.60 |

(Determined by G. Davidson, Witteck Developments Inc.)

Estimated grade of tungsten concentrate, based on lead removal

$$= \frac{0.16\ WO_3}{t\ conc} \times \frac{0.27\ W}{WO_3} \times \frac{303.693\ FeWO_4}{183.85\ W} \times \frac{0.60\ PbS}{1.0\ FeWO_4}$$

$$= \frac{0.0428\ t\ PbS}{t\ conc}$$

Average tonnage to bulk sulfide circuit = 1.7887 mt/hr gravity concentrate.

For 95% removal of lead in bulk sulfide circuit, the weight loss

$$= 1.7887\ mt/hr \times \frac{0.0428\ PbS}{t\ conc} \times 0.95$$

= 0.0727 mt/hr PbS

## 2.6 AMMONIUM PARATUNGSTATE HYDROMET

Assumption 7:

Extraction of tungsten in a caustic leach was determined to be 96% and the loss of tungsten in the purification step was 5.4-7.7% of net

tungsten in leach feed (DNR Process Studies Report, No. F-16, May, 1988). The major loss of tungsten is associated with the leach and purification unit operations.

Average tonnage of tungsten concentrate to hydromet
= 1.7887 mt/hr - 0.0727 mt/hr
= 1.7160 mt/hr solids

Tungsten in hydromet feed = (0.1755 + 0.1107) - 0.0055
= 0.2807 mt/hr $WO_3$

Amount of tungsten lost in leach residue = 0.2807 mt/hr (1 - 0.96)
= 0.0112 mt/hr $WO_3$

For a leach residue of 0.57% $WO_3$, total solids = $\frac{0.0112 \text{ mt/hr } WO_3}{0.0057}$

= 1.969 mt/hr solids

Amount of tungsten to purification = 0.2807 - 0.0112 mt/hr $WO_3$
= 0.2695 mt/hr $WO_3$

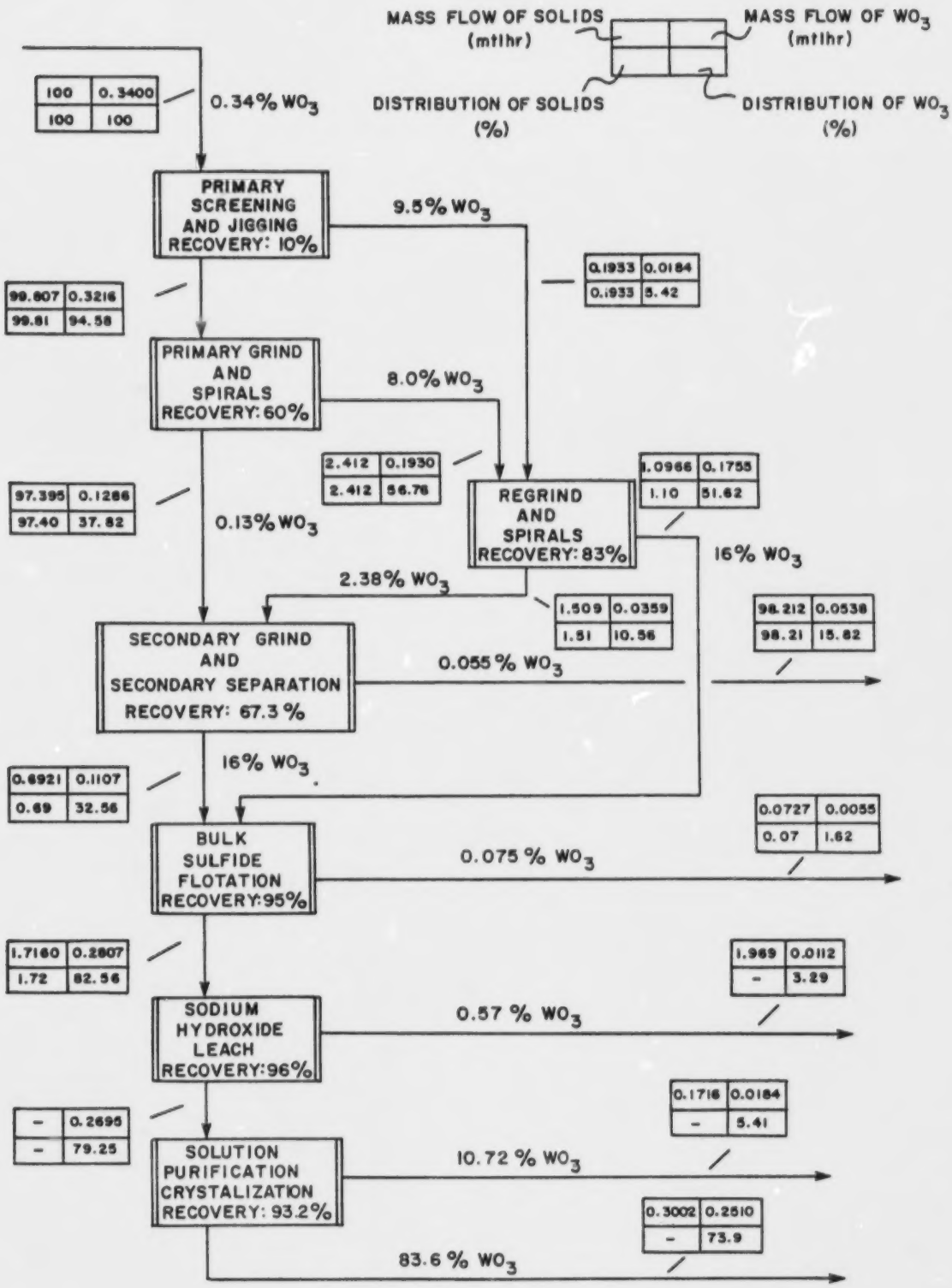
Amount of tungsten lost in purification = 0.2807 x 0.0655
= 0.0184 mt/hr $WO_3$

Total tungsten to solvent extraction and crystallization
= 0.2695 - 0.0184
= 0.2501 mt/hr $WO_3$

Overall hydromet recovery = $\frac{0.2510 \text{ mt/hr } WO_3}{0.2807 \text{ mt/hr } WO_3}$

= 89.4%

# MASS BALANCE SUMMARY



## 3.0 EQUIPMENT SIZING FOR GRINDING/GRAVITY SEPARATION

### 3.1 PRIMARY AND SECONDARY SCREENS

From Norberg Process Machinery Reference Manual, 11-24, p. 30.

Theoretical area required:

$$= \frac{\text{total undersize throughput mt/hr} \times 1.1023}{A \times B \times C \times D \times E \times F \times G \times H \times 10.69} = m^2$$

| SCREEN | OPENING SIZE | UNDERSIZE THROUGHPUT (MT/HR) | CAPACITY SIZING FACTORS | | | | | | | | AREA $M^2$ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | A | B | C | D | E | F | G | H | |
| Primary | 4 mesh<br>4.76 mm (0.187 in) | 48 | 0.89 | 0.98 | 1.07 | 1.00 | 2.50 | 1.50 | 0.90 | 1.00 | 1.57 |
| Secondary | 28 mesh<br>595 um (0.0234 in) | 5 | 0.18 | 0.44 | 1.07 | 1.00 | 1.00 | 1.50 | 0.90 | 1.00 | 4.506 |

a. Primary Undersize: For crushed ore at -16 mm, the amount less than 4 mesh is approximately 30-50%, depending on the crushing circuit configuration. Use maximum tonnage for design as outlined in 2.1.

b. Secondary Undersize: For crushed ore, the amount of material -28 mesh is approximately 5-10%. Use maximum tonnage for secondary screen sizing.

Specific gravity of material 2.5 $mt/m^3$ = 157.29 $lb/ft^3$

Primary screen = 1.57 $m^2$ (16.87 $ft^2$) 2' x 8' single deck screen
Secondary screen = 4.51 $m^2$ (48.2 $ft^2$) 4' x 12 ' single deck screen

### 3.2 PRIMARY ROD MILL

Estimated tonnage to rod mill

From primary screen = 57 mt/hr
From secondary screen = 43 mt/hr
From rougher spiral = 10 mt/hr (10% spiral feed)
Total tonnage = 110 mt/hr (121 st/hr)

Estimated particle size of feed = 595 - 16,000 um
F = size having 80% passing = 12,000 um

Estimated particle size of discharge = 1000 - 1200 um
P = size having 80% passing = 841 um

For a work index of 13.5, the power consumption is expressed by:

$$kWh/st = \frac{10(13.5)}{(841)^{0.5}} - \frac{10(13.5)}{(12,000)^{0.5}}$$

$$= 3.42 \text{ kWh/st}$$

For wet grind open circuit correction factor 1.25

Top Horse Power = 3.42 x 1.25 x 121 x 1.34
= 693 Hp (554-693 Hp)

For mill power at 35-45% volumetric loading, mill size range is 10.0 ft x 13.5 ft to 10.5 ft x 14.5 ft (based on Allis-Chalmers Mill Sizing Data). Maximum rod charge is 79 mt.

### 3.3 TERTIARY AND REGRIND DSM SCREENS (SIEVE BENDS)

Design basis: For 2' width, maximum tonnage 15-20 mt/hr

Tertiary screen = 4 screen[2] (14 mesh)

Regrind screen = 1 screen (32 mesh)

### 3.4 JIG

Design maximum throughput for a 24" x 36" jig = 15-20 mt/hr

Number of jigs = 4

### 3.5 SPIRALS (PRIMARY AND REGRIND CIRCUIT)

Estimated tonnage to primary grind rougher = 115 mt/hr (max)
58 spirals

cleaner = 5-10 mt/hr
5 spirals

Estimated tonnage to regrind rougher = 3-5 mt/hr
3 spirals

cleaner = 2-4 mt/hr
2 spirals

NOTE: Spiral sizing based on a capacity of 2 mt/hr of dry solids.

### 3.6 REGRIND BALL MILL

Estimated tonnage = 5 mt/hr maximum (5.51 st/hr)

Estimated particle size of feed = 500-1000 um (predominantly from primary grind circuit)

F = size having 80% passing = 800 um

Estimated particle size of product = 100-300 um
P = size having 80% passing = 275 um

For a work index of 13.5, the power consumption is expressed by:

$$kWh/st = \frac{10(13.5)}{(275)^{0.5}} - \frac{10(13.5)}{(800)^{0.5}}$$

$$= 3.36 \text{ kWh/st}$$

For a wet grind, open circuit, the correction factor is 1.25

Maximum horsepower = 3.36 x 5.51 x 1.25 x 1.34
= 31 Hp

For a mill power of 19-47 Hp at 35-45% volumetric loading, the mill size ranges from 4 ft x 4 ft to 5 ft x 5 ft.

### 3.7 CYCLONE SIZING FOR PRIMARY, SECONDARY AND TERTIARY SEPARATION CIRCUITS

(Based on Krebs Cyclone Performance Chart, Bulletin No. 21-130)

Primary: Tonnage = 98.4 mt/hr Flowrate = 269 $m^3$/hr 30% solids

Cut point = 200-250 um

Estimated size of cyclone = 26-30 in diameter

Select 26 in to operate at 5 psi (capacity 272 $m^3$/hr)

Number of cyclones = 1

Secondary: Tonnage: Feed tonnage to secondary cyclone is dependent on the ball mill performance and mass balance of primary cyclone.

Assumption: Base the mass balance on a mill discharge in the range of minus 300-500 um, therefore, the feed to the secondary cyclone contains 40-60% minus 100 um material (average of 50%).

Tonnage of secondary cyclone overflow
= feed to primary - conc produced by coarse and intermediate separation
= 98.9 - 0.60
= 98.3 mt/hr

Feed to secondary cyclone = $\frac{98.3}{0.50}$ = 196.6 mt/hr

Flowrate = 537 $m^3$/hr 30% solids

Cut point = 50-110 um

Estimated size of cyclone = 15-20 in

Select 15 in to operate at 10 psi (capacity 90 $m^3$/hr)

Number of cyclones = 6

Tertiary Tonnage = 97.8 mt/hr Flowrate = 430.5 $m^3$/hr 20% solids

Cut point = 25-35 um

Estimated size of cylcone = 4 in

Operating at 10-15 psi, capacity = 4.5-8.0 $m^3$/hr

Number of cyclones = 54-96

### 3.8 SECONDARY BALL MILL

Estimation of tonnage based on secondary cyclone feed.

Assumption: Overflow tonnage from primary cyclone is 20-30% of feed (25% average)

Mill tonnage = 196.6 - 0.25(98.9)
= 172 mt/hr (189.6 st/hr)

Recirculating load = $\frac{172 \text{ mt/hr}}{97.8 \text{ mt/hr}} \times 100\%$

= 175% (200%)

Mill size based on 97.8 mt/hr

Estimated particle size of feed = 100-1000 um
F = size having 80% passing = 600 um

Estimated particle size of product = -300 um
P = size having 80% passing = 150 um

For a work index of 13.5, the estimated power consumption is expressed by:

$$kWh/st = \frac{10(13.5)}{(150)^{0.5}} - \frac{10(13.5)}{(600)^{0.5}}$$

= 5.51 kWh

For wet grind closed circuit, correction factor is 1.0

Maximum horsepower = 5.51 kWh/st x 98.7 x 1.34
= 729 Hp

Select 11.5 ft x 11 ft ball mill (mill power is 674-782 for 35-45% volumetric loading). Maximum ball charge is 60 mt.

### 3.9 REICHERT CONE SIZING

Based on an average capacity of 50 mt/hr.

Primary cyclone underflow = 98.9 - 0.25 (98.9)
= 74.2 mt/hr

Rougher estimated tonnage (max) 79 mt/hr, 2 units required
Cleaner estimated tonnage (max) 5 mt/hr, 1 unit required

### 3.10 SECONDARY SEPARATION TABLES

NOTE: Double deck tables with a capacity of 1 mt/table deck (2 mt/unit).

| | | |
|---|---|---|
| Coarse Tables: | Rougher: | Tonnage = 196.6 - 97.8<br>98.8 mt/hr |
| | | Number of units = 50 |
| | Cleaner: | Tonnage (maximum) = 6 mt/hr |
| | | Number of units = 3 |
| Intermediate Tables: | Rougher: | Tonnage = 97.8 - 0.15(97.8)<br>= 83 mt/hr |
| | Assumption: | Tonnage of tertiary cyclone overflow is 10-20% of feed tonnage (average 15%). |
| | | Number of units = 42 |
| | Cleaner: | Tonnage (maximum) = 4 mt/hr |
| | | Number of units = 2 |
| Fine Tables: | Rougher: | Tonnage = 0.15(97.8)<br>14.67 |
| | | Number of units = 7-10* |
| | Cleaner | Tonnage (maximum) = 1 mt/hr |
| | | Number of units = 1 |

* depending on the operating conditions

## 4.0 TUNGSTEN UPGRADING/APT HYDROMET

### 4.1 BULK SULFIDE FLOTATION

From Mass Balance (Section 2.0)

Tonnage primary separation circuit = 0.69 mt/hr

Tonnage from B/S flotation = 0.073 mt/hr

A. **GRAVITY CONCENTRATE THICKENER AND TUNGSTEN CONCENTRATE THICKENER**

Feed % solid = 16-33%

Unit area requirement = 0.0615 $m^2$/mt/day (design basis)

$$\text{Total area for settling} = \frac{0.0615 \times 0.69 \times 24}{0.16}$$

$$= 6.36\ m^2$$

$$\text{Volumetric flow to thickener} = \frac{0.69\ mt/hr}{0.16} \times \frac{m^3}{1.136\ mt}$$

$$= 3.795\ m^3/hr$$

Rise velocity of thickener = 0.5967 m/hr

Diameter of thickener = 2.8 m (9.33 ft)

$$\text{Rate horsepower for heavy duty thickener} = \frac{8000\ ft \cdot lb \times 2\ rpm}{5250 \times 0.60\ efficiency}$$

$$= 5\ Hp$$

B. **GRAVITY CONCENTRATE HOLDING TANK AND TUNGSTEN CONCENTRATE HOLDING TANK**

Feed % solid = 50-60%

Residence time requirements = 5 hr

$$\text{Flowrate} = \frac{0.69\ mt/hr}{0.55} \times \frac{m^3}{1.702\ mt}$$

$$\text{Size of tank} = 0.73 \times 5$$
$$= 3.7\ m^3$$

$$\text{Agitator horsepower} = 3.7\ m^3 \times \frac{2.5\ Hp}{3.79\ m^3} \times 1.10\ \text{factor}$$

$$= 3.0\ Hp$$

C. CONDITIONING TANK

Percentage solids (feed to bulk sulfide flotation) = 45%

Residence time requirements = 30 min

$$\text{Flowrate} = \frac{0.69 \text{ mt/hr}}{0.45} \times \frac{\text{m}^3}{1.509 \text{ mt}}$$

$$= 1.016 \text{ m}^3/\text{hr}$$

$$\text{Size of tank} = 1.016 \text{ m}^3/\text{hr} \times 0.5 \text{ hr}$$
$$= 0.5 \text{ m}^3$$

$$\text{Agitator horsepower} = 0.5 \text{ m}^3 \times \frac{2.5 \text{ Hp}}{3.79 \text{ m}^3} \times 1.10 \text{ factor}$$

$$= 0.36 \text{ Hp } (\tfrac{1}{2} \text{ Hp})$$

D. REGRIND CYCLONE

Estimated flowrate = 1.5 $m^3$/hr

Cut point = 25-65 um

Cyclone size = 3-4 in (select 3 in)

Number of cyclones = 1 (1.59 $m^3$/hr at 4 psi)

E. REGRIND MILL

Mill tonnage (maximum) = 0.85 mt/hr (0.9367 st/hr)

Feed size range = 0-300 um
F size having 80% passing = 225 um

Product size range = 0-100 um
P size having 80% passing = 75 um

For a work index of 13.5, the estimated power consumption i expressed by:

$$\text{kWh/st} = \frac{10(13.5)}{(75)^{0.5}} - \frac{10(13.5)}{(225)^{0.5}}$$

$$= 6.58 \text{ kWh/st}$$

Maximum horsepower = 6.58 kWh/st x 0.9367 x 1.34
= 8.25 Hp (average = 10 Hp)

Select 3 ft x 3 ft mill having 7-9 Hp for 35-45% volumetric loading ball charge 0.87 mt.

## 4.2 CAUSTIC DIGESTION OF TUNGSTEN CONCENTRATES

NOTE: Commercial operating conditions for the reaction of caustic with tungsten concentrates is based on a 10 molar concentration of caustic and the consumption is 1.5 stoichiometric requirements. If a leaching unit operation were employed using a 10 molar concentration, the consumption of caustic would be high. Therefore, the sizing of this unit operation will be based on digestion.

The design basis for gravity concentrate will be based on the following test program assays:

| ASSAY | RATIO TO $WO_3$ | |
|---|---|---|
| 16% $WO_3$ | 1 | |
| 0.056% P | 0.0035 | ()estimate |
| 0.027% Cu | 0.0017 | |
| 0.103% Mo | 0.0064 | |
| 7.66% As | 0.4788 | |
| (3.40%) Pb | 0.2125 | |
| (25%) Si | 1.56 | |

The stoichiometry of caustic addition is based on the following reaction:

$$FeWO_4 + 2NaOH = Fe(OH)_2 + Na_2WO_4$$

Commercial conditions for caustic digestion are:

a. 1.5 times stoichiometric requirements.
b. caustic digestion done at 10 molar NaOH using 50% caustic solution.

Hydrometallurgical sizing is based on a throughput of concentrate containing (minimum grade) 16% $WO_3$.

Throughput = 0.69 mt/hr x 24 hr/day
= 16.560 mt/day

## A. TUNGSTEN CONCENTRATE THICKENER AND FILTER SIZING

Thickener size is similar to gravity concentrate thickener.

Assumption: filter sizing is based on a filtration rate of 200 Kg dry solid/hr/$m^2$

$$\text{Maximum filter area required} = \frac{0.69 \text{ mt/hr} \times 1000 \text{ kg/mt}}{200 \text{ kg dry solid/hr/m}^2}$$

$$= 3.5 \text{ m}^2$$

## B. STOICHIOMETRIC REQUIREMENTS OF CAUSTIC

Consumption of NaOH = 16.56 t/day x 0.16 t $WO_3$/t conc
x 303.69 t $FeWO_4$/231.85 t $WO_3$
x 1 mole $FeWO_4$/303.69 t $FeWO_4$
x 2 mole NaOH/mole $FeWO_4$
x 39.98 t NaOH/mole NaOH

= 0.9138 t/day NaOH

Based on stoichiometric requirements, consumption of NaOH
= 1.5 x 0.9138 t/day
= 1.3707 mt/day NaOH
= (0.0571 mt/hr)

Consumption of caustic solution (50% w/w)
= 1.3707 mt NaOH/day / 0.50 mt NaOH/mt soln
= 2.7414 mt/day NaOH

Volumetric flowrate of caustic
= 2.7414 mt/day x 1 day/24hrs x $m^3$/1.48 mt
= 0.0772 $m^3$ /hr caustic solution
(0.1143 t/hr)

## C. VOLUME OF WATER ADDED TO DIGESTION

Concentration of caustic to be 10 moles NaOH/L soln

Total volume of soln = 0.0571 mt/hr x 1000 kg/t
/(39.98 kg/mole x 10 kg mole/$m^3$)
= 0.1429 $m^3$/hr

Volume of water = 0.1429 $m^3$/hr - 0.0772 $m^3$/hr caustic soln
= 0.0657 $m^3$/hr

Weight percentage solids in digestion = $\frac{0.69\ t \times 100\%}{0.69\ t + 0.1143 + 0.0657}$

= 79%

D. SIZING OF CAUSTIC DIGESTION VESSEL

Total volumetric throughput of slurry = $\frac{0.69\ mt/hr}{0.79} \times \frac{m^3}{2.597\ mt}$

= 0.3363 $m^3/hr$

Total residence time required for digestion = 4 hr

Total volume with correction factor for short-circuiting
= 0.3363 $m^3/hr$ x 4 hr x 1.20
= 1.614 $m^3$ (56.95 $ft^3$)

For design, the average mixing capacity is 60 $ft^3$

Based on Marion Mixers, volume of reactor is 84 $ft^3$, (Catalog No. 188).

Standard horsepower for mixing = 0.8734 t/hr x 4 hr x 1000 kg/t
x 0.0165 Hp/kg x 1.10
= 63.4 Hp

Average powered reg'd = 65 Hp

E. HEATING REQUIREMENTS FOR CAUSTIC DIGESTION

Minimum temperature of slurry = 15°C

Maximum temperature of bake = 100°C

Change in slurry temperature = 85°C

Heat capacity of slurry = 0.40 to 0.60 Kcal/kg°C

Steam from boiler at 100 psi with 80% heating efficiency
= 528 Kcal/kg

Average steam flow reg'd for heat up to reaction temperature
= $\frac{0.3363\ m^3/hr \times 2597\ kg/m^3 \times 85°C \times 0.60\ Kcal/Kg°C}{528\ Kcal/kg\ steam}$

= 84.4 Kg/hr

Assumption: Steam required to offset heat loss from digester is based on 5 to 10% (average 7.5%) of steam requirement to heat cold feed to reaction temperature.

Total steam demand = 84.4 kg/hr + 0.075 (84.4 kg/hr)
= 90.7 kg/hr steam at 100 psi

## 4.3 WATER LEACH CIRCUIT

The water leach circuit is designed to continue the leaching action of caustic in a well agitated tank and to avoid the poisoning of the reaction by the formation of iron hydroxide coatings on particles.

The water addition to the water leach circuit is based on a 4.5-5.5 molar NaOH solution to allow for the dissolution of sodium tungstate.

Maximum water addition is based on a 4.5 molar leach.

Total volume of soln = 0.0571 mt/hr x 1000 kg/t
/(39.98 kg/mole x 4.5 kg mole/$m^3$)
= 0.3174 $m^3$/hr

Volume of soln in digestion = 0.1429 $m^3$/hr + 0.0657 $m^3$/hr
= 0.2086 $m^3$/hr

Water addition to leach = 0.3174 - 0.2086 $m^3$/hr
= 0.1088 $m^3$/hr

### A. SIZING OF WATER LEACH

Total residence time required for leach is 2 hours.

Total volume of leach circuit = (0.3363 + 0.1088 $m^3$/hr) x 2
= 0.89 $m^3$

Total volume with correction factor for short-circuiting (1.25)
= 1.11 $m^3$

Volume of each leach tank (3 tanks) = 0.37 $m^3$ (97.9 USGal)

Average capacity of leach tanks = 0.38 $m^3$ (100 USGal)

Agitator horsepower for each tank = 0.38 $m^3$ x 0.66 Hp/$m^3$
x 1.1 factor
= 0.27 Hp (¼ Hp)

**B. HEAT REQUIREMENTS FOR WATER LEACH CIRCUIT**

Minimum temperature of slurry = 90°C

Maximum temperature of leach = 100°C

Average flowrate of slurry = 0.4451 $m^3$/hr

Steam from a boiler at 100 psi with 80% heating efficiency
= 528 Kcal/kg

Average steam req'd for heat up = (0.4451 x 2189 x 10 x 0.7)/528
= 12.9 kg/hr

Total steam demand including heat loss = 12.9 + 0.075(12.9)(2)
= 14.8 kg/hr

**C. LEACH RESIDUE HEAT EXCHANGER**

To avoid flashing of leach liquor in filtrate receivers:

Maximum temperature of leach residue = 100°C
Maximum temperature of filtrate feed = 55°C

Based on maximum slurry flow and water

MCp T water = MCp T slurry

M (1 Kcal/kg°C)(35 - 20) = 978.8 kg(0.7 Kcal/kg°C)(100 - 55)

Mass flow of cooling water = 2055 kg/hr
= 2.0 $m^3$/hr (7.5 g/m)

## 4.4 LEACH RESIDUE - LIQUOR SEPARATION

Weight loss of solids in leach circuit is 30-35% of feed solid weight.

Holding tanks and repulp tanks operate at 60% solids for optimum washing.

**A. HOLDING AND REPULP TANK SIZING**

Residence time in each tank = 5 hr

Maximum throughput of solids = 0.69 mt/hr x (1 - 0.30)
= 0.4830 mt/hr

Amount of solution from leach = 0.1143 + 0.0657 + 0.1088 mt/hr
= 0.2888 mt/hr

Percentage solid in slurry = 62.6% w/w (1.885 mt/$m^3$)

Volume of each tank = $\frac{(0.4830 + 0.2888)\text{mt/hr} \times 5\text{ hr}}{1.885\text{ mt/m}^3}$

= 2.047 $m^3$ (540 USGal)

Agitator horsepower = 2.047 $m^3$ x 0.66 Hp/$m^3$ x 1.1 factor
= 1.48 Hp (1.5 Hp)

B. **DRUM FILTER**

Assumption: Sizing is based on a filtration rate of 100-200 kg dry solids/hr/$m^2$

Maximum filter area required = $\frac{0.4830\text{ mt/hr} \times 1000\text{ kg/mt}}{100\text{ kg/hr/m}^2}$

= 4.83 $m^2$

C. **FILTRATE FLOWRATE**

Amount of solution retained in filtrate cake at 15% moisture
= 0.0852 $m^3$/hr

Flow of filtrate from each filtration step = 0.2888 - 0.0852 $m^3$/hr
= 0.2036 $m^3$/hr

Total flow of filtrate to purification = 0.4071 $m^3$/hr (1.5 gpm)

D. **CONSUMPTION OF CAUSTIC FOR WASH SOLUTION**

Amount of wash solution required = 0.2036 $m^3$/hr

A 10% caustic solution used for wash.

Amount of caustic consumed = 0.2036 $m^3$/hr x 1100 kg/$m^3$ x 0.10
= 22.39 kg/hr

Consumption of caustic solution = 44.79 kg/hr
= 1.075 mt/day of 50% soln

## 4.5 SOLUTION PURIFICATION

NOTE: Magnesium chloride has been reported in the literature for the removal of silica etc. and with ammonia hydroxide under the following conditions:

| | |
|---|---|
| $MgCl_2$, 80-90°C, pH 10.5-11.5 | Si, Fe, Al, Nb, Ti |
| $MgCl_2$ + $NH_4OH$, 70-80°C, pH 8-9 | P, As |

The addition of these cementation reagents is based on the stoichiometry of the following reaction:

$$Na_2HXO_4 + MgCl_2 + NH_4OH = Mg(NH_4)XO_4 + 2NaCl + H_2O$$

where x = P, As, Mo, Si

Amount of x in the concentrate = 0.056% P + 7.66% As
+ 0.103% Mo + 23% Si
= 30.8% X

Assumption: Calculations are based on an extraction of 10% of the impurities in the caustic leach
: Let X represent silica, because of the higher concentration in the concentrate.

NOTE: test results indicate a limited extraction of impurities (silica was not measured).

Amount of $Na_2HSiO_4$ formed = 0.69 mt/hr x 0.308 Si x 0.10 ext
x 139.1 t $Na_2HSiO_4$/28.09 t Si
= 0.1052 t/hr $Na_2HSiO_4$
= 105.2 kg/hr or 0.7563 kg/mole/hr

### A. MAGNESIUM CHLORIDE CONSUMPTION

Amount of $MgCl_2$ = 0.7563 $\frac{kg/mole}{hr}$ x 95.211 kg $MgCl_2$/mole

= 72.01 kg/hr $MgCl_2$

Based on concentrate = 0.1043 t $MgCl_2$/t concentrate *

* Consumption used in test program.

**B. AMMONIUM HYDROXIDE CONSUMPTION**

Amount of $NH_4OH$ = 0.7563 kg mole/hr x 35.0 kg $NH_4OH$/mole
= 26.47 kg/hr $NH_4OH$

Bulk solution of ammonium hydroxide contains 30% $NH_3$ (61.74% $NH_4OH$)

Amount of ammonium hydroxide soln

$$= \frac{26.47 \text{ kg/hr } NH_4OH \times 17.0 \text{ Kg } NH_3}{0.30 \ NH_3/\text{soln} \times 35.0 \text{ kg } NH_4OH}$$

= 42.9 kg/hr of ammonium solution

Based on concentrate = 0.0621 t $NH_3$ soln/t concentrate

Volumetric flow of solution = 0.69 mt/hr x 0.0621 $\frac{\text{t soln}}{\text{t conc}}$ $\frac{\text{m}^3}{0.89 \text{ mt}}$

= 0.0481 $m^3$/hr (0.17 gpm)

**C. MAGNESIUM CHLORIDE BINS**

Bulk density of approximately 1.0 t/$m^3$ purchased as $MgCl_2 \cdot 6H_2O$

Volume requirements for 30 day storage

$$= 72 \text{ kg/hr } MgCl_2 \times \frac{303.31 \ MgCl_2 \cdot 6H_2O}{95.31 \ MgCl_2} \times \frac{\text{m}^3}{1000 \text{ kg}} \times 720 \text{ hr}$$

= 164.97 $m^3$

For a bin with 60° cone:

$$\text{Single bin diameter} = \left[\frac{0.65(82.49)}{0.9527}\right]^{0.333}$$

= 3.80 m (12.46 ft)

Total height = 7.88 m (25.80 ft)

**D. CEMENTATION/SOLUTION PURIFICATION TANKS**

Residence time in purification circuit = 2 hr

Throughput of solution = 0.4071 $m^3$/hr

Volume of each tank (3) = 0.4071 $m^3$/hr x 2 hr/3
= 0.27 $m^3$ (71.6 USGal)

Agitator horsepower = 0.27 $m^3$/hr x 0.30 Hp/$m^3$ x 1.1 factor
= 0.089 Hp (1/8 Hp)

Purification holding tank residence time = 5 hr

Volume of each tank = 0.4071 $m^3$/hr x 5 hr
= 2.035 $m^3$ (537 USGal)

Agitator horsepower = 2.035 $m^3$ x 0.30 Hp/$m^3$ x 1.1 factor
= 0.67 Hp (3/4 Hp)

**E. HEAT REQUIREMENTS FOR PURIFICATION TANKS**

Minimum temperature of solution = 35°C
(heat loss from filtration)

Maximum temperature of solution = 95°C

Average flowrate of solution = 0.4071 $m^3$/hr

Steam from a boiler at 100 psi with 80% heating efficiency
= 528 Kcal/kg

Average steam requirement for heat up
= (0.4071 x 1050 x 60 x 1.0)/528
= 48.57

Total steam demand for both circuits including heat loss
= 2(43.57 + 3.5(0.075)(48.57))
= 122.7 kg/hr steam

**F. ADDITION OF SULFURIC ACID (PH MODIFIER)**

For a pH of 8, the addition of sulfuric acid is based on 0.10 L of 20% $H_2SO_4$ per kg of concentrate.

Consumption of sulfuric acid = 0.69 mt conc/hr x 1000 kg
x 0.10 L/kg x 0.20

= 13.8 L/hr (of 98% $H_2SO_4$
sp. gr. 1.8)
= 24.8 kg/hr

## 4.6 SOLVENT EXTRACTION

The solvent extraction circuit design is based on:

Extraction

Concentration of $WO_3^{-2}$ = 150 g/L ($kg/m^3$)

Organic/aqueous ratio = 1.3

Maximum contact time = 10 min

Maximum setting time = 5 min

Number of stages = 2

Strip

Organic/aqueous ratio = 3.5

Maximum contact time = 10 min

Maximum setting time = 5 min

Number of stages = 1

Organic Solution Composition

12% Tricaprlylamine
12% Tributyl Phosphate
76% Kerosene

### A. SODIUM TUNGSTATE SOLUTION (STS) HOLDING TANK

Residence time of holding tank = 5 hr

Throughput of solution based on concentration of $WO_3$
= 150 $kg/m^3$ (g/L)

Maximum concentration in purified solution
= 0.69 mt/hr x 1000 kg/mt x 0.16 $WO_3$/0.4071 $m^3$/hr
= 211.2 $kg/m^3$ $WO_3$

Amount of solution required for design concentration
= 0.69 (1000)(0.16)/150
= 0.7360 $m^3$/hr

Amount of process water added to STS holding tank
= 0.3289 $m^3$/hr

Volume of tank = 0.7360 $m^3$/hr x 5 hr
= 3.68 $m^3$

Agitator horsepower = 3.68 $m^3$ x 0.30 Hp/$m^3$ x 1.1 factor
= 1.21 Hp (1½ Hp)

B. EXTRACTION MIXER SETTLER

Throughput of aqueous solution = 0.7360 $m^3$/hr (2.69 g/m)

Organic recycle = 0.7360 $m^3$/hr x 1.3
= 0.9568 $m^3$/hr (3.51 g/m)

Sizing of mixing cell = (0.9568 $m^3$/hr + 0.7360 $m^3$/hr)
x 1 hr/60 min x 10 min

= 0.2821 $m^3$/hr (62 gal)

Sizing of settling chamber = 0.2821 $m^3$ x 10 min/5 min
= 0.1411 $m^3$ (31 gal)

C. STRIPPING MIXER SETTLER

Amount of ammonium hydroxide is based on the stoichiometry of the following:

$Na_2WO_4 + 2NH_4OH = (NH_4)_2WO_4 + 2NaOH$

Amount of $Na_2WO_4$ = 0.69 mt/hr x 0.16 $WO_3$ x 1.267 $Na_2WO_4/WO_3$
= 0.1399 mt/hr

Amount of $NH_4OH$ required

$$= 0.1399 \text{ mt/hr} \times \frac{\text{1 mole}}{293.93 \text{ t } Na_2WO_4} \times \frac{2 \text{ mole } NH_4OH}{\text{mole } Na_2WO_4} \times \frac{35 \text{ t } NH_4OH}{\text{mole}}$$

= 0.033 mt $NH_4OH$/hr

Throughput of organic solution = 0.9568 $m^3$/hr

Throughput of aqueous strip solution = 0.9568/3.5 $m^3$/hr
= 0.2734 $m^3$/hr

$$\text{Concentration of } NH_4OH \text{ solution} = \frac{0.033 \text{ mt/}NH_4OH \times 1000 \text{ kg/mt}}{0.2734}$$

= 120.7 kg/$m^3$ (3.5 molar)

Stock solution of $NH_4OH$ = 61.74% $NH_4OH$

Consumption of $NH_4OH$ solution

$$= \frac{120.7 \text{ kg } NH_4OH}{m^3} \times 0.2734 \text{ m}^3/\text{hr} \times \frac{\text{kg soln}}{0.6174 \text{ kg } NH_4OH}$$

= 53.45 kg/hr

D. **SULFATION TANK**

Consumption of sulfuric acid based on stoichiometry of the following reaction:

$$2NaOH + H_2SO_4 = Na_2SO_4 + 2H_2O$$

Consumption of NaOH in digestion = 0.0571 mt/hr (section 4.2B)

$$\text{Amount of sulfuric acid} = 0.0571 \frac{\text{mt}}{\text{hr}} \times \frac{\text{mole NaOH}}{39.89 \text{ mt}} \times \frac{1 \text{ mole } H_2SO_4}{2 \text{ mole NaOH}}$$

$$= 0.070 \text{ mt/hr } (0.35 \text{ m}^3/\text{hr})$$

Residence time for sulfation = 15 min

Flow of organic = 0.9568 $m^3/hr$

Volume of tank = $(0.9568 + 0.035) m^3/hr \times 1 hr/60 min \times 15 min$
= 0.25 $m^3$

Agitator horsepower = 0.25 $m^3$ x 0.30 $Hp/m^3$ x 1.1 factor
= 0.08 Hp (1/8 Hp)

E. **WASH RAFFINATE HOLDING TANK**

Residence time = 5 hr

Flowrate of aqueous strip solution = 0.2734 $m^3/hr$

Tank volume = 0.2734 x 5
= 1.37 $m^3$

## 4.7 CRYSTALLIZER

Solution flow to crystallizer = 0.2734 $m^3/hr$ (1.0 g/m)

Maximum production rate of APT
= 100 mt/hr ore x 0.0034 $WO_3$ x 0.80 rec x $\frac{0.793\ W}{WO_3}$ x $\frac{APT}{0.7043\ W}$

= 0.3063 mt/hr APT

Concentration of APT = 1,120 kg/$m^3$ (272 kg $WO_3$/hr)

**A. SELECTION OF CRYSTALLIZER**

Draft-Tube-Baffle (DTB) Evaporator-Crystallizer

- mechanical circulation to increase nucleation
- prevents short-circuiting of solids in supersaturation zone
- heated solution is mixed with recirculating slurry, may reduce scale formation by dissolving scale with unsaturated solution
- fine destruction so product range is 6-50 mesh

**B. MASS BALANCE**

Feed = 0.2734 $m^3$/hr x 1.05 mt/$m^3$ x 1000 kg/mt
= 287 kg/hr

Wash water from product filter
= 306.3 kg/hr APT x 0.12 kg wash/kg APT
= 36.756 kg/hr

Amount of liquid with product filter cake (10% moisture)

$$= \left[ \frac{0.3063 \text{ mt/hr}}{0.90} - 0.3063 \right] \times 1000$$

= 34.03 kg/hr

Maximum evaporation rate assuming no purge required
= 287 kg/hr + 36.756 - 34.03
= 289.7 kg/hr

**C. HEAT BALANCE (APPROXIMATION)**

Sensible heat = 289 kg/hr x 20°C x 1 Kcal/kg°C
= 5,740 Kcal/hr

Heat of crystallization = 306.2 kg/hr x 57 Kcal/kg
= 17,453 Kcal/hr

Latent heat of vapor = 287 kg/hr x 573 Kcal/kg
= -164,541 Kcal/kg

Steam load = 141,348 Kcal/hr (0.07 mm BTU/hr)

Steam from a boiler at 100 psi with 80% heating efficiency
= 528 Kcal/kg

Average steam requirement = 141,348/528
= 267.7 kg/hr

Total steam demand including heat loss (75%)
= 267.7 + (0.075)(267.7)
= 287.7 kg/hr (634 lb/hr)

D. **VOLUME OF CRYSTALLIZER**

Sizing calculations are based on a retention time of 5 hr

Volume = 0.2734 $m^3$/hr x 5 hr
= 1.36 $m^3$ (300 gal)

Total volume including evaporation space = 2(1.36 $m^3$)
= 2.72 $m^3$

Estimated diameter = 0.61 m

## 4.8 PRODUCT DEWATERING

Assumption: Sizing of filter is based on a filtration rate of 50 kg dry solids/hr/$m^2$

Maximum filtration area = 306.3 kg/hr / 50 kg dry solid/hr/$m^2$
= 6.2 $m^2$

# APPENDIX B

## PREFEASIBILITY OF AMMONIUM PARATUNGSTATE PRODUCTION AT THE MOUNT PLEASANT MINE

## ORDER OF MAGNITUDE CAPITAL COST ESTIMATE

## JANUARY, 1989

## 1.0 BASIS FOR COST ESTIMATES

- Purchase cost for major equipment expenditures is based on supplier budget (order of magnitude) quotes.
- In most cases, transportation cost is a factored cost of the purchased equipment cost.
- Installation costs are based on equipment supplier recommendations or factored based on construction experience.
- Electrical, instrumentation and piping costs are factored based on the installation of comminution, beneficiation and hydrometallurgical operations.
- Total capital cost does not include the construction of a building for metallurgical operations, and civil works is included in installation.

NOTE: For each section, the suggested suppliers are indicated in brackets.

## 2.0 MECHANICAL EQUIPMENT COST ESTIMATES

### 2.1 GRINDING EQUIPMENT
(Allis Chalmers, Sala Machinery Works and Dominion Engineering)

| | | |
|---|---|---|
| A. | Rod Mill - 3.2 m dia x 4.57 m<br>700 Hp motor liners and charge<br>(maximum rod charge = 68 mt) | 890,000 |
| B. | Ball Mill - 3.51 m dia x 3.35 m<br>750 Hp motor and liners<br>(maximum ball charge = 61 mt) | 924,000 |
| C. | Ball Mill - 1.52 m dia x 1.52 m<br>35 Hp motor and liners<br>(maximum ball charge = 4.7 mt) | 123,500 |
| D. | Ball Mill - 0.91 m dia x 0.91 m<br>10 Hp motor and liners<br>(maximum ball charge = 0.87 mt) | 97,000 |
| | Purchase Equipment Cost | 2,034,000 |
| | Transportation Cost | 100,000 |
| | Installation Cost | 915,000 |

2.2 **SCREENS**
**(Derrick Screens, Rexnord and Sala Machine Works)**

| | | |
|---|---|---|
| A. | Vibrating Screen - 4' x 8', 5 Hp | 22,100 |
| B. | Vibrating Screen - 6' x 12', 10 Hp | 27,600 |
| C. | DSM Screen (14 mesh) | 22,000 |
| D. | DSM Screen (32 mesh) | 22,000 |
| | Purchase Equipment Cost | 93,700 |
| | Transportation Cost | 3,000 |
| | Installation Cost | 18,000 |

2.3 **GRAVITY SEPARATION EQUIPMENT**
**(Carpco, Joy)**

| | | |
|---|---|---|
| A. | Jig - 24" x 36", 2 Hp (1 unit) | 15,000 |
| B. | Spirals (2 mt/hr) (68 units) | 306,000 |
| C. | Reichert Cones (50 mt/hr) (3 units) | 277,000 |
| D. | Double Deck Tables, 1 mt/deck, 2 Hp (108 units) | 5,100,000 |
| | Purchase Equipment Cost | 5,699,000 |
| | Transportation Cost | 171,000 |
| | Installation Cost | 1,417,000 |

2.4 **CYCLONES**
**(Mosley, Krebs)**

| | | |
|---|---|---|
| A. | Cyclone - Rubber lined, CS Body, 26" dia (1 unit) | 10,000 |
| B. | Cyclone - Rubber lined, CS Body, 15" dia distributor (6 units) | 40,000 |
| C. | Cyclone - Rubber lined, CS Body, 4" dia distributor (96 units) | 423,000 |
| D. | Cyclone - Rubber lined, CS Body, 3" dia (1 unit) | 4,000 |
| | Purchase Equipment Cost | 477,000 |
| | Transportation Cost | 20,000 |
| | Installation Cost | 190,000 |

**2.5 DIGESTER**
**(Marion Mixer)**

| | |
|---|---|
| A. Continuous Mixer - 1.6 $m^3$ 316SS Pressure Rated Jacket 65 Hp motor (1 unit) | 96,000 |
| Purchase Equipment Cost | 96,000 |
| Transportation Cost | 4,000 |
| Installation Cost | 20,000 |

**2.6 CRYSTALLIZER**
**(Whiting Equipment)**

| | |
|---|---|
| A. Crystallizer Vessel - 2.72 $m^3$, 316SS Agitator (2 Hp) Recycle Pump (2 Hp) Heat Exchanger | 250,000 |
| Purchase Equipment Cost | 250,000 |
| Transportation Cost | 10,000 |
| Installation Cost | 250,000 |

**2.7 SOLVENT EXTRACTION**
**(Estimate is based on beryllium, REF USBM IC 9143, Krebs, Hazen Quinn)**

| | |
|---|---|
| A. Mixer-settler, storage tanks, mixer, and pumps - FRP Feedrate of pregnant solution at 15 L/min | 160,000 |
| Purchase Equipment Cost | 160,000 |
| Transportation Cost | 8,000 |
| Installation Cost | 50,000 |

## 2.8 CENTRIFUGAL SLURRY PUMPS
(Worthington, Gauvreau)

All pumps rubber lined, gland water bushing, belt driven

| | | |
|---|---|---|
| A. | 1½" x 1½", 2 Hp motor (1 unit) | 2,500 |
| B. | 3" x 3", 35 Hp motor (1 unit) | 4,500 |
| C. | 4-5" x 3-5", 30-35 Hp motor (5 units) | 27,500 |
| D. | 6-8" x 5-6", 50 Hp motor (9 units) | 76,500 |
| E. | Vertical Pump, 1 Hp motor (5 units) | 12,500 |
| | Purchase Equipment Cost | 123,500 |
| | Transportation Cost | 5,000 |
| | Installation Cost | 49,500 |

## 2.9 CENTRIFUGAL CHEMICAL SEALESS PUMPS
(Wilfley, Warmen)

All pumps stainless steel, direct drive, centrifugally sealed

| | | |
|---|---|---|
| A. | 1.5" x 1.5", 2 Hp (2 units) | 10,600 |
| B. | 1.0" x 1.0", 0.5-1 Hp (5 units) | 15,000 |
| C. | 0.5" x 0.5", 0.5 Hp (2 units) | 3,000 |
| | Purchase Equipment Cost | 28,600 |
| | Transportation Cost | 1,000 |
| | Installation Cost | 11,400 |

## 2.10 AIR DIAPHRAM PUMPS
(Dorr Oliver)

| | | |
|---|---|---|
| A. | 2" Slurry Pump for THK.U/F Air Operated (4 units) | 12,000 |
| | Purchase Equipment Cost | 12,000 |
| | Transportation Cost | 500 |
| | Installation Cost | 5,000 |

**2.11 SOLID-LIQUID SEPARATION, FILTERS**
**(Denver, Dorr Oliver)**

| | |
|---|---|
| A. Belt Filter - 3.5 $m^2$ with Drive and Filtrate Receivers (1 unit) | 88,000 |
| B. Pressure Filter - (2 units) | 20,000 |
| C. Drum Filter - 5 $m^2$ with Drive and Filtrate Receivers (1 unit) | 60,000 |
| D. Drum Filter - 6.2 $m^2$ with Drive and Filtrate Receivers (1 unit) | 72,000 |
| Purchase Equipment Cost | 240,000 |
| Transportation Cost | 7,000 |
| Installation Cost | 60,000 |

**2.12 SOLID-LIQUID SEPARATION, SETTLING VESSELS**
**(Denver, Dorr Oliver)**

| | |
|---|---|
| A. Thickener - 2.8 m dia, 5 Hp drive Rake Control (2 units) | 80,000 |
| Purchase Equipment Cost | 80,000 |
| Transportation Cost | 4,000 |
| Installation Cost | 36,000 |

**2.13 TANKS AND AGITATORS**
**(Greey Lightning, Teckstar Plastics)**

Tanks fabricated from polethylene, with cover, agitator and agitator mount, and flanged connections.

| | | |
|---|---|---|
| A. | 3.7 $m^3$, 3.0 Hp (2 units) | 15,600 |
| B. | 0.5 $m^3$, 0.5 Hp (1 unit) | 1,200 |
| C. | 0.4 $m^3$, 1.0 Hp (3 units with steam coils) | 7,000 |
| D. | 2.1 $m^3$, 1.5 Hp (3 units) | 12,500 |
| E. | 0.3 $m^3$, 0.5 Hp (6 units with steam coils) | 8,000 |
| F. | 2.04 $m^3$, 0.75 Hp (2 units) | 6,000 |
| G. | 3.7 $m^3$, 1.25 Hp (1 unit) | 5,000 |
| H. | 0.25 $m^3$, 0.25 Hp | 600 |
| I. | 1.37 $m^3$ | 1,100 |
| | Purchase Equipment Cost | 57,000 |
| | Transportation Cost | 2,000 |
| | Installation Cost | 20,000 |

**2.14 STORAGE BINS AND FEEDERS**

| | | |
|---|---|---|
| A. | Carbon Steel Bin - 85 $m^3$<br>72 kg/hr Vibration Feeder<br>(2 units) | 68,000 |
| | Purchase Equipment Cost | 68,000 |
| | Transportation Cost | 2,000 |
| | Installation Cost | 14,000 |

**2.15 ELECTRICAL EQUIPMENT**

A. Based on purchase cost of mechanical equipment, cost includes switch gear, transformers, motor control centre.

| | |
|---|---|
| Purchase Equipment Cost | 350,000 |
| Transportation Cost | 10,000 |
| Installation Cost | 53,000 |

**2.16 INSTRUMENTATION EQUIPMENT**
**(Leeds and Northrup)**

A. Based on an integrated controller, with 8 controller files with backup logic power supply, graphic trending, 2 screens, and 18 m of optical highway. 470,000

| | |
|---|---|
| Purchase Equipment Cost | 470,000 |
| Transportation Cost | 5,000 |
| Installation Cost | 70,500 |

**3.0 ELECTR**

Includes cable tray, conduit, lighting, wire and cable

| | |
|---|---|
| Bulk Material | 954,000 |
| Labour | 477,000 |

**4.0 INSTRUMENTATION**

Includes sensors and transmitters for analog and digital input/output, support tray, wire and cable, localized instrumentation and gauges.

| | |
|---|---|
| Bulk Material | 867,000 |
| Labour | 433,000 |

**5.0 PIPING**

Includes piping, launders, plumbing, and inline valves, unions, gas hets, hangers, insulation, etc.

| | |
|---|---|
| Bulk Material | 1,300,000 |
| Labour | 780,000 |
| Total | 2,080,000 |

# APPENDIX C

## PREFEASIBILITY OF AMMONIUM PARATUNGSTATE PRODUCTION AT THE MOUNT PLEASANT MINE

## ORDER OF MAGNITUDE OPERATING COST ESTIMATE

## JANUARY, 1989

## 1.0 BASIS FOR COST ESTIMATES

- Wages for employees are based on standard industrial practice within New Brunswick and overtime, burden and fringe benefits are are factored.

- Reagent costs are based on published prices in the Chemical Marketing Reporter (copyright by Schnell Publishing Company Inc.).

- Operating and maintenance costs are based on estimated requirements using standard industrial practice.

- Operating cost for crushing, product drying and packaging, and tailing treatment are factored costs.

## 2.0 LABOUR COST FOR GRINDING, GRAVITY AND HYDROMET OPERATIONS

NOTE: Wage rates are based on New Brunswick industrial practice.

| | TITLE | DOLLARS PER YEAR | NO. MEN | TOTAL DOLLARS PER YEAR |
|---|---|---|---|---|
| Supervision Staff | Mill Superintendent | 60,000 | 1 | 60,000 |
| | General Mill Foreman | 45,000 | 1 | 45,000 |
| | Metallurgist | 40,000 | 2 | 80,000 |
| | Chemist/Assayers | 35,000 | 3 | 105,000 |
| | Lab Technicians | 25,000 | 4 | 100,000 |
| | Shift Foreman | 40,000 | 4 | 160,000 |
| Hourly-Rated Staff | (42 Hours/Week) | | | |
| | Control Rm. Operator | 40,000 | 4 | 160,000 |
| | Grinding Operator | 35,000 | 4 | 140,000 |
| | Gravity Operator | 35,000 | 4 | 140,000 |
| | Hydromet Operator | 35,000 | 4 | 140,000 |
| | Shift Labourer | 25,000 | 12 | 300,000 |
| | Gravity Assistant | 25,000 | 4 | 100,000 |
| | Sampler | 25,000 | 4 | 100,000 |
| Hourly-Rated Staff | (40 Hour Basis) | | | |
| | Electrical-Mech Superintendent | 50,000 | 1 | 50,000 |
| | Electricians | 35,000 | 2 | 70,000 |
| | Millwrights | 35,000 | 4 | 140,000 |
| | Mechanics | 35,000 | 3 | 105,000 |
| | Machinist, Welder | 35,000 | 2 | 70,000 |
| | Shop Foreman | 45,000 | 2 | 90,000 |
| | Instrumentation Technician | 30,000 | 2 | 60,000 |
| | Labourers, Greaser | 25,000 | 4 | 100,000 |
| | TOTAL STAFF: | | 56 | |
| | TOTAL ANNUAL SALARY | | | 2,315,000 |
| | OVERTIME (10%) | | | 232,000 |
| | BURDEN/FRINGE BENEFITS (32%) | | | 740,000 |
| | NET TOTAL ANNUAL SALARY | | | 3,287,000 |

## 3.0 REAGENTS CONSUMED

### A. FLOTATION REAGENTS

Price of flotation reagents FOB mine site = $0.29/kg

Maximum consumption = 0.20 kg/mt

Net annual expenditure = 0.2 kg/mt x 2400 mt/day x 365 day/yr
x $0.29/kg
= $50,800/yr

### B. SODIUM HYDROXIDE (50% W/W SOLUTION)

Price = $270/mt FOB Montreal
Shipping = $ 82/mt
Cost on site = $352/mt

= 0.1143 mt/hr caustic leach
= 0.044 mt/hr wash solution
Total consumption = 0.1583 mt/hr

Total annual expenditure = $352/mt x 1,386.7 mt/yr
= $488,100/yr

### C. AMMONIUM HYDROXIDE (AQUEOUS AMMONIA 29.4% $NH_3$)

Price = $378/mt FOB Montreal
Shipping = $ 82/mt
Cost on site = $460/mt

Purification = 0.043 mt/hr
Hydrolysis (SX Strip) = 0.054 mt/hr
Total consumption = 0.097 mt/hr

Total expenditure = $460/mt x 849.72 mt/yr
= $390,900/yr

D. **ORGANIC REPLACEMENT**

Replacement cost for organic used in solvent extraction is dependant on solvent loss, availability and degradation of the chemicals. By industrial experience, the loss of amine is 0.009% and 0.07% for kerosene.

Makeup cost for amine = 0.77 mt x $3,125/mt
= $2,406

Makeup cost for kerosene = 5.8 $m^3$ x 1000 $L/m^3$ x $0.44/L
= $2,563

Total makeup cost = $5,000 (rounded off)

E. **SULFURIC ACID (98%-100%)**

Price = $115/mt FOB Montreal
Shipping = $ 82/mt
Cost on Site = $197/mt

24.8 kg/hr for pH modification in purification
70 kg/hr for sulfation of organic in solvent extraction.

Total consumption = 0.095 mt/hr

Total annual expenditure = $197/mt x 830 mt/yr
= $163,600/yr

F. **MAGNESIUM CHLORIDE, HYDROUS (99%)**

Price = $330/mt FOB Montreal
Shipping = $ 82/mt
Cost on Site = $412/mt

72 kg/hr for purification

Total annual expenditure = $412/mt x 630 mt/yr
= $259,900/yr

## 4.0 OPERATING AND MAINTENANCE SUPPLIES

| | | |
|---|---|---|
| A. | Grinding Media (0.40 kg steel/mt ore) | $105,000/yr |
| B. | Replacement Liners for All Mills (once per year) | 200,000/yr |
| C. | Replacement Screens | 15,000/yr |
| D. | Maintenance Supplies (8% of equipment purchase cost) | 694,000/yr |
| E. | Filter Cloths | 20,000/yr |
| F. | Safety Supplies | 10,000/yr |
| G. | Assay and Metallurgical Supplies | 206,000/yr |
| | TOTAL COST | $1,250,000/yr |

## 5.0 ELECTRICAL POWER

| AREA | INSTALLED HORSEPOWER | HP USED |
|---|---|---|
| Grinding | 1495 @ 75% | 1121 |
| Gravity Separation, Screens | 618 @ 75% | 464 |
| APT Hydromet | 99 @ 75% | 74 |
| Building Lighting, Exhaust Fans, Etc. (5% of equipment power) | | 83 |
| TOTAL POWER | | 1742 |

Total cost at (¢4.24/Kw·Hr) = 1742 Hp x 0.7320 Kw/Hp x 8760 hr/yr
x $0.0424/Kw·hr
= $473,600/yr

## 6.0 WATER

Total consumption of water based on 50% solid in tails = 98 $m^3$/hr

Total annual cost at ($0.013/$m^3$) = $11,300/yr

## 7.0 STEAM

| STEAM DEMAND | DISTRIBUTION |
|---|---|
| Caustic Digestion = 91 kg/hr | 17.6% |
| Water Leach = 15 kg/hr | 2.9% |
| Solution Purification = 123 kg/hr | 23.8% |
| Crystallization = 288 kg/hr | 55.7% |

Total Process Steam Demand = 517 kg/hr

Plant Heating Steam Demand (4 months) = 1,200 kg/hr

Total annual cost at ($0.0110/kg) = 88,400

NOTE: About 56% of the steam consumed is used for process heating.

## 8.0 CRUSHING

NOTE: Value of feedstock is based on an operating cost for crushing of ore (reference to USBM IC 9143).

| For 2,400 mt/day | $1984 | $1988 |
|---|---|---|
| Equipment Operating Cost/Day | 455 | 492 |
| Labour Cost/Day | 1,970 | 2,127 |
| Supplies/Day | 218 | 235 |
| Total Cost/day | | 2,854 |

Annual Cost = $1,041,700

## 9.0 PRODUCT DRYING AND PACKAGING

NOTE: Based on reference to USBM IC 9143 and packaging in 30 gal drums.

| For 10 mt/day: Concentrate Drying Cost | $ 1984 | $1988 |
|---|---|---|
| Equipment Operating Cost/Day | 141 | 152 |
| Labour Cost/Day | 196 | 212 |
| Supplies/Day | 91 | 98 |
| Total Drying Coast Per Day | | 462 |
| Cost for Packaging Based on $30/m Packaged/Day | | 300 |
| Total Cost Per Day Drying & Packaging | | 762 |

Annual Cost = $271,300

## 10.0 TAILING TREATMENT

ASSUMPTION: Treatment cost for most operations is $0.90/$m^3$ tailing (Cost includes tailing pond management)

Total tailing discharge based on 98 mt/hr solids and 50% solids.

Annual Cost = 150 $m^3$/hr x 8,760 hr/yr x $0.90 $m^3$
= 1,182,600

## 11.0 CONTRACTORS

| | | |
|---|---|---|
| A. | Maintenance on Special Jobs (2% Equip. Cost) | 173,000 |
| B. | Technical and Financial Audits (0.17% of Revenue) | 40,000 |
| C. | Equipment Service Calls (1% of Equip. Cost) | 87,000 |
| | TOTAL ANNUAL COST | 300,000 |

## 12.0 INSURANCE

For loss due to fire etc., rate is approximately $0.35/100 capital.

Annual payment = 0.0035 ($15,788,100)
= $55,300

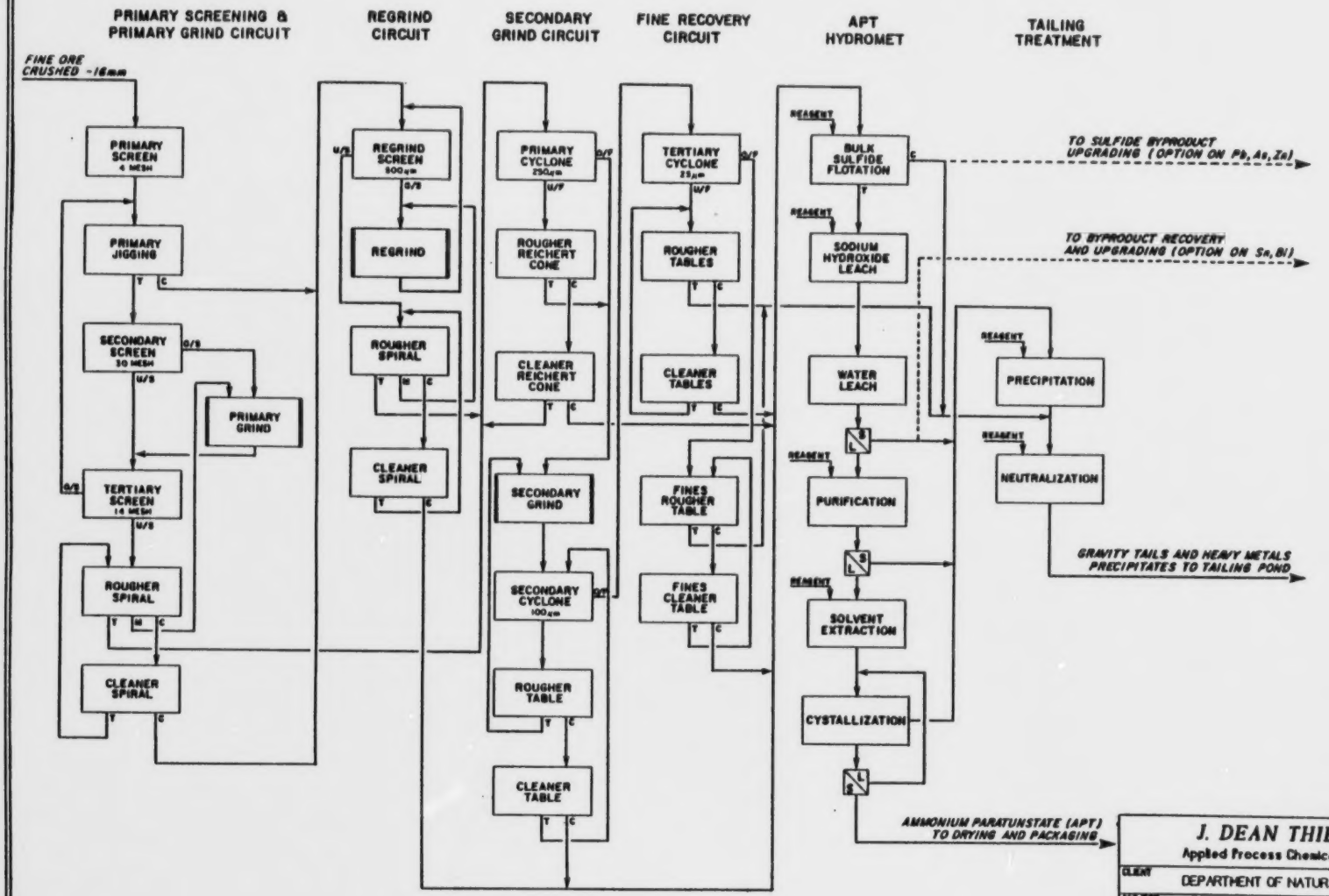
# APPENDIX D

## PREFEASIBILITY OF AMMONIUM PARATUNGSTATE PRODUCTION AT THE MOUNT PLEASANT MINE

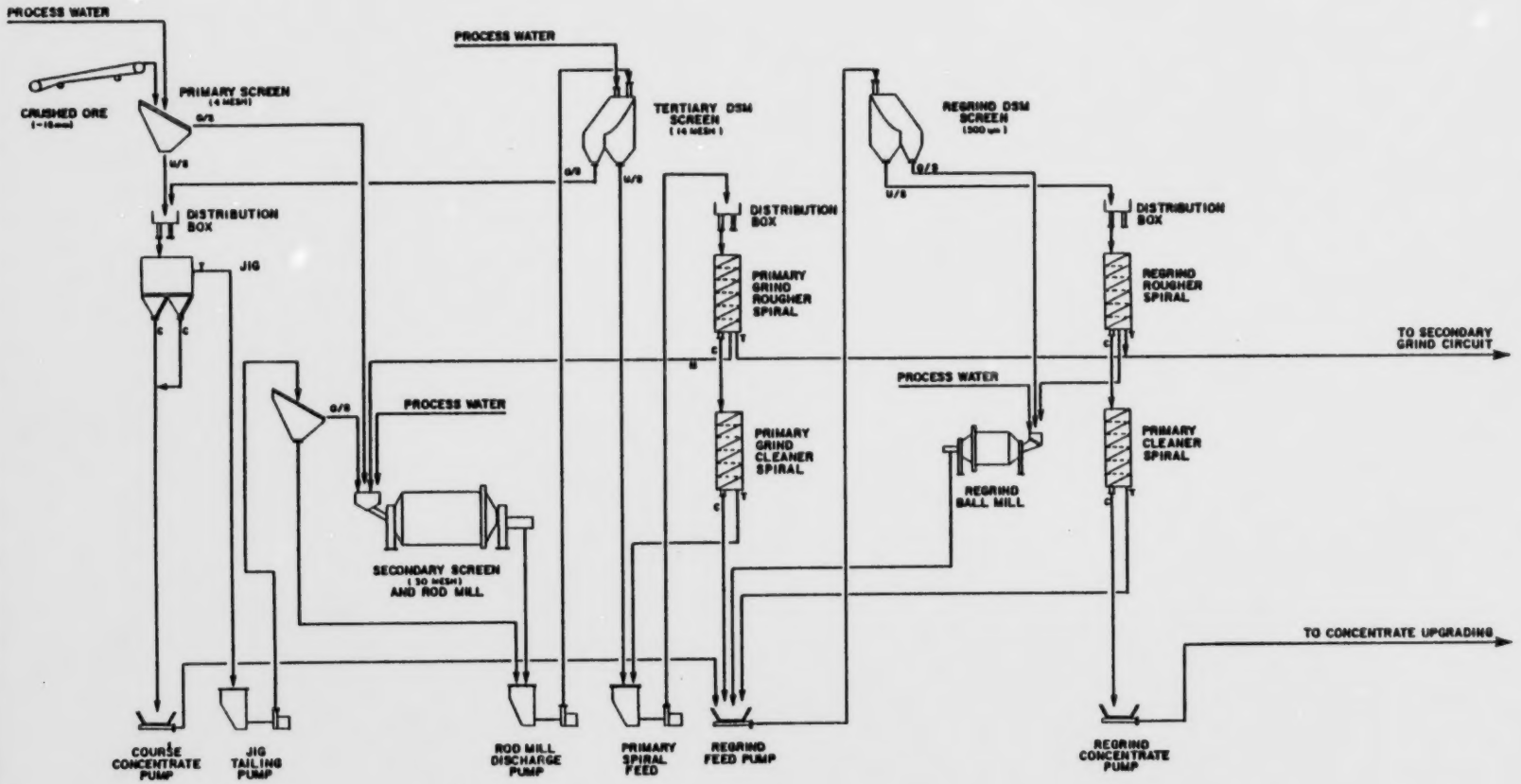
## PROCESS FLOW DIAGRAMS

JANUARY, 1989

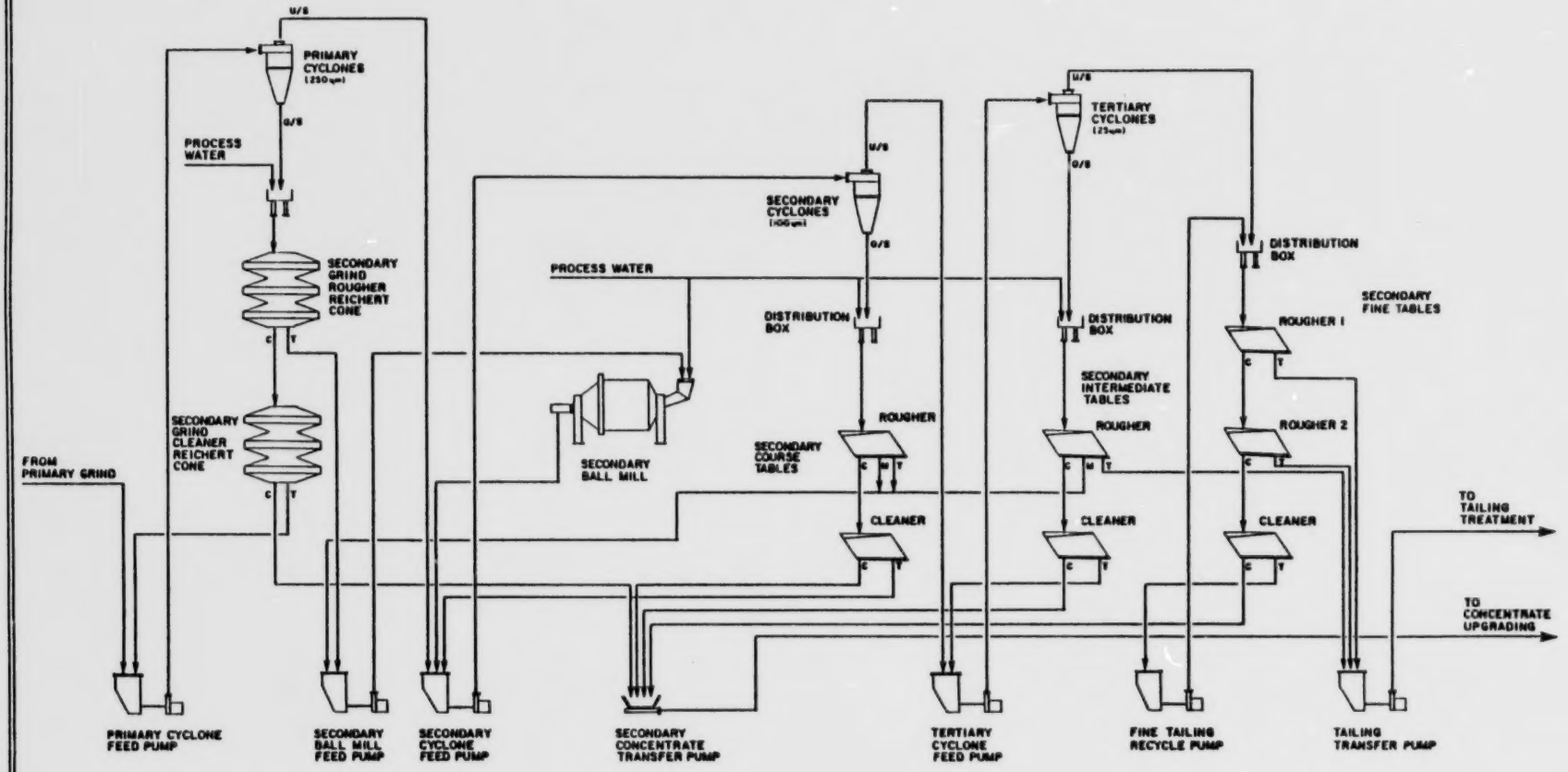
| DRAWING NUMBER | TITLE |
|---|---|
| 0100-01-01 | Process Block Diagram: Gravity and APT Hydromet |
| 2100-01-02 | Process Flow Diagram: Primary Grind and Regrind |
| 3100-01-03 | Process Flow Diagram: Secondary Grind |
| 4100-01-04 | Process Flow Diagram: Tungsten Upgrading and Leaching |
| 4100-02-05 | Process Flow Diagram: Purification and Crystallization |



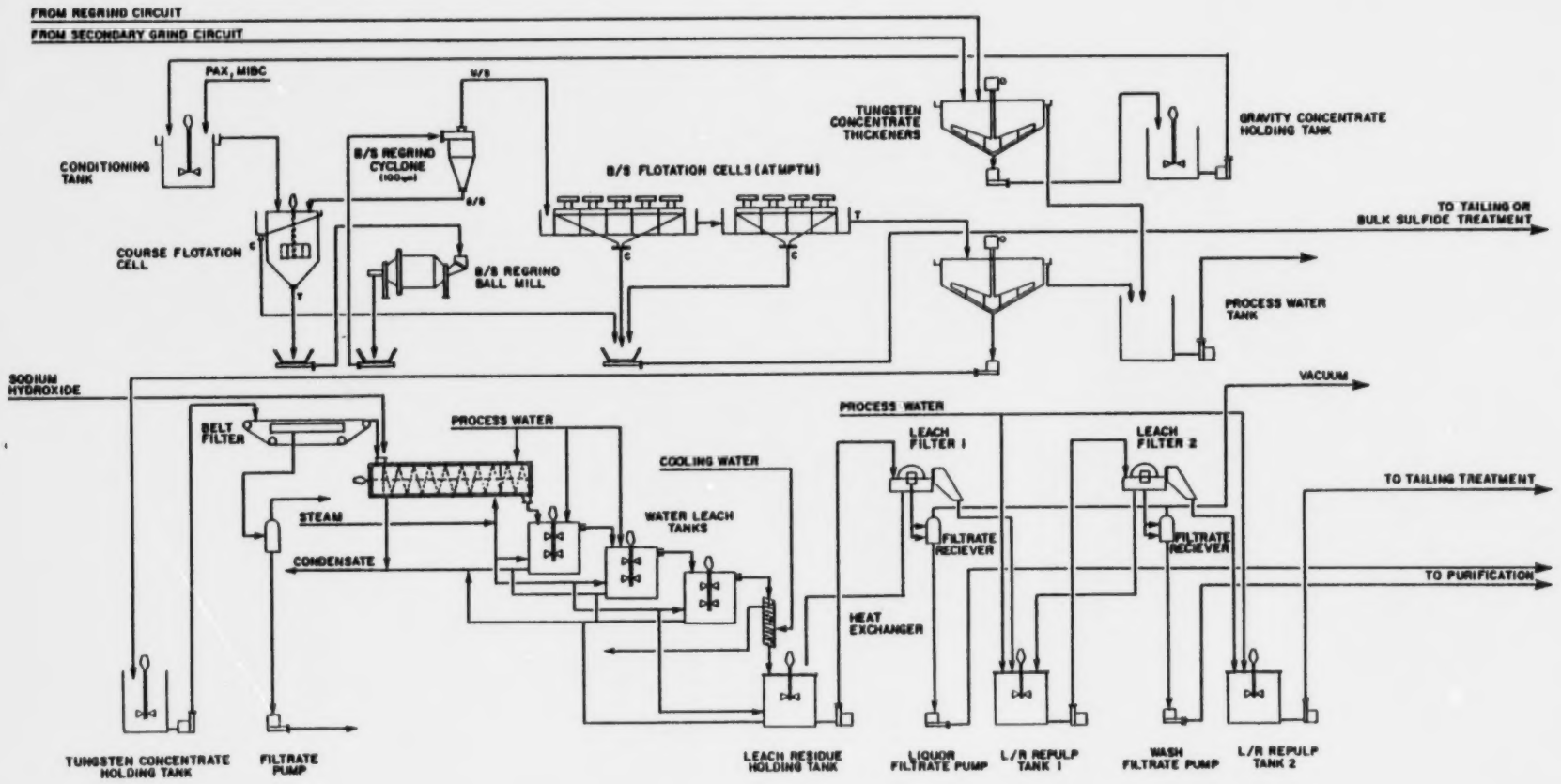

| | | | |
|---|---|---|---|
| J. DEAN THIBAULT. P. ENG. Applied Process Chemical Engineering Consultant | | | |
| CLIENT | DEPARTMENT OF NATURAL RESOURCES & ENERGY | | |
| PROJECT | PREFEASIBILITY OF APT PRODUCTION AT THE MOUNT PLEASANT MINE | | |
| TITLE | Block Diagram : Gravity Seperation / Hydromet | | |
| DESIGN | J.D. Thibault | DATE 01.01.89 | REVISION NUMBER 0 |
| DRAWN | R. Phillips | DATE 14.03.89 | JOB NUMBER 8808 |
| APPROVED | | DATE | DRAWING NUMBER 0100-01-01 |




| | | | |
|---|---|---|---|
| *J. DEAN THIBAULT. P. ENG.* Applied Process Chemical Engineering Consultant | | | |
| CLIENT | DEPARTMENT OF NATURAL RESOURCES & ENERGY | | |
| PROJECT | PREFEASIBILITY OF APT PRODUCTION AT THE MOUNT PLEASANT MINE | | |
| TITLE | Process Flow Diagram : Primary Grind / Regrind | | |
| DESIGN J.D. Thibault | DATE 01.01.89 | REVISION NUMBER | 0 |
| DRAWN R. Phillips | DATE 14.03.89 | JOB NUMBER | 8808 |
| APPROVED | DATE | DRAWING NUMBER | 2100-01-02 |




| | | | |
|---|---|---|---|
| *J. DEAN THIBAULT. P. ENG.* Applied Process Chemical Engineering Consultant | | | |
| CLIENT | DEPARTMENT OF NATURAL RESOURCES & ENERGY | | |
| PROJECT | PREFEASIBILITY OF APT PRODUCTION AT THE MOUNT PLEASANT MINE | | |
| TITLE | Process Flow Diagram : Secondary Grind | | |
| DESIGN | J.D. Thibault | DATE 01.01.89 | REVISION NUMBER 0 |
| DRAWN | R. Phillips | DATE 14.03.89 | JOB NUMBER 8808 |
| APPROVED | | DATE | DRAWING NUMBER 3100-01-03 |

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| CLIENT | DEPARTMENT OF NATURAL RESOURCES & ENERGY | | |
| PROJECT | PREFEASIBILITY OF APT PRODUCTION AT THE MOUNT PLEASANT MINE | | |
| TITLE | Flow Diagram : Tungsten Upgrading & Leaching | | |
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| DRAWN R. Phillips | DATE 14.03.89 | JOB NUMBER | 8808 |
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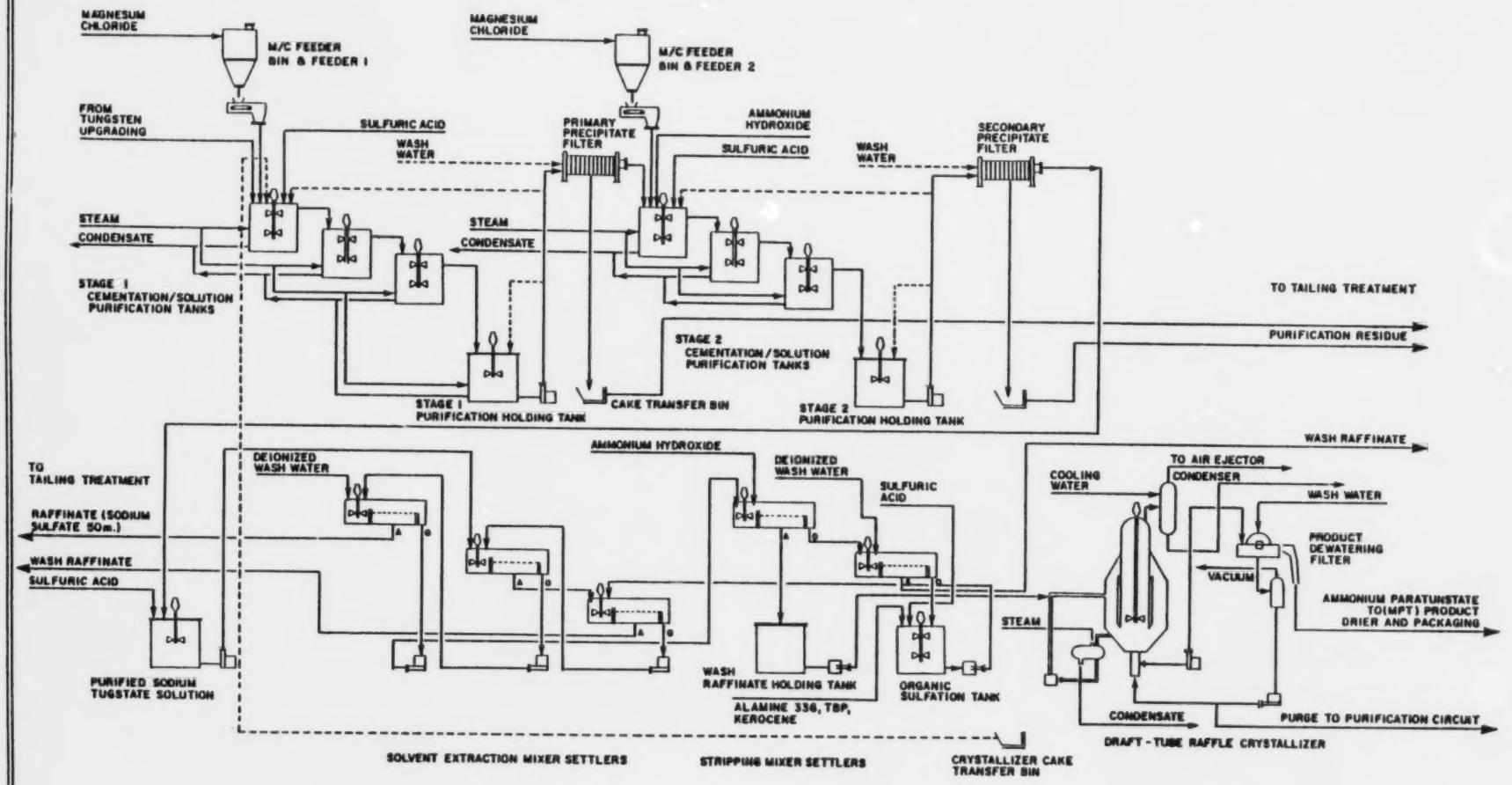


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| TITLE | Process Flow Diagram : Purification & Crystallization | | |
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## 1.0 SOMMAIRE

La fermeture de la mine de Mount Pleasant découle de plusieurs facteurs, y compris les limites techniques de la technologie du procédé et une baisse de la valeur sur le marché des concentrés de tungstène. Des tentatives de la dernière chance, en vue de résoudre les problèmes de qualité et de production du produit, avec l'aide d'une subvention du ministère de l'Expansion industrielle régionale (MEIR), se sont soldées par un échec par suite du coût excessif de la production. Depuis la fermeture de la mine, Lac Minerals Ltd. a effectué une étude de faisabilité de la production de concentrés d'étain à partir d'une zone de minerai adjacente tandis que le ministère des Ressources naturelles et de l'Énergie s'est intéressé, de son côté, aux options de remplacement pour l'utilisation du minerai de tungstène.

Dans les conditions actuelles du marché, le prix du tungstène demeure encore le principal facteur économique de la valeur du minerai de la zone de la Tour d'observation nord. En plus du tungstène, plusieurs concepts variés de développement en puissance pour la production de sous-produits (tels que le fluorure de calcium, le molybdène, les sulfures en vrac de cuivre et de plomb ainsi que le zinc, le bismuth et l'étain) ont été déterminés. Cette étude est centrée sur les options envisageables d'utilisation de dérivés du tungstène et met l'accent sur la production de paratungstate d'ammonium (PTA) à partir de minerai de tungstène.

Ce rapport présente l'évaluation de préfaisabilité de la production de PTA à Mount Pleasant, réalisée à partir d'un programme pilote sur une échelle réduite effectuée par le ministère des Ressources naturelles et de l'Énergie. Les principales conclusions de cette étude peuvent se résumer comme suit :

- La méthode proposée d'enrichissement primaire des minerais, par l'intégration du broyage et de la séparation par gravité, est un procédé unique à l'industrie du tungstène et devrait s'avérer plus efficace que la technologie par flottation. La technologie du procédé de production de PTA par gravité à partir de concentrés a été mise au point par divers fabricants. Cependant, l'élaboration de la technologie du procédé pour les minerais de Mount Pleasant devrait être réalisée en tenant plus compte de la qualité de minerai particulier, en raison des diverses concentrations d'impuretés qu'il renferme.

- Les coûts en capital de la construction d'une usine capable de produire 2 235 tonnes de PTA par année dans le cadre des installations actuelles à Mount Pleasant devraient être d'environ 28 millions de dollars canadiens (janvier 1989). Cette estimation comprend l'achat et la mise en place de l'équipement de transformation ainsi que les coûts indirects du broyage, de la séparation par gravité et de l'usine chimique de PTA. Le devis estimatif des coûts d'exploitation du traitement du minerai, après extraction, se chiffre à 13,50 $ la tonne de minerai soit 4,50 $ le kilo de PTA.

- Le marché du PTA est très concurrentiel ainsi le cassage des prix pratiqué par les Chinois est un des principaux éléments de dissuasion pour le développement de l'industrie nord-américaine du tungstène. Pour un rendement de 10 % de l'investissement sur 10 ans (avant impôt) la valeur du PTA devrait dépasser 150 $ la tonne métrique de trioxide de tungstène contenue dans le concentré afin de rendre viable la production de PTA à partir du minerai de Mount Pleasant. C'est là deux fois la valeur actuellement cotée sur le marché.

- Les prix actuels de la poudre et du carbure de tungstène ne sont pas touchés par les pratiques commerciales chinoises. La production de carbure de tungstène et de métaux durs à partir de ce minérai permet d'envisager des revenus environ cinq fois supérieurs à ceux que l'on retirerait du PTA pour deux fois le montant de l'investissement en capital requis. Le rendement en puissance de l'investissement dans le cas d'une installation de production entièrement intégrée à Mount Pleasant est économiquement plus attrayant que la seule production de concentré ou de PTA ou de ces seuls deux produits.

- Étant donné que la production de PTA n'est qu'une étape intermédiaire du procédé, le développement d'une installation de traitement entièrement intégrée impliquerait l'élaboration d'une technologie du procédé. Tout ajout de technologie additionnelle pour le traitement et la production de poudre métallique et de carbure entrerait dans le cadre normal des pratiques industrielles. De ce fait, le coût de la mise au point de la technologie à Mount Pleasant serait équivalent à celui de l'élaboration d'une technologie de production de PTA.